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SPENT LIQUOR RECOVERY, 1977.

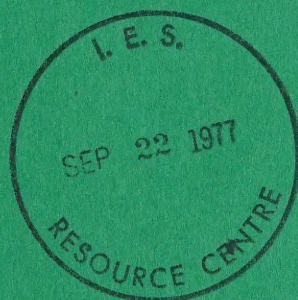
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Technology and Associated Cost for Sulphite Pulping Spent Liquor Recovery



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Report EPS 3-WP-77-8

Water Pollution Control Directorate
June 1977

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TECHNOLOGY AND ASSOCIATED COST FOR SULPHITE
PULPING SPENT LIQUOR RECOVERY

by

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for the

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FISHERIES AND ENVIRONMENT CANADA

Report No. EPS 3-WP-77-8
June 1977

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This report has been reviewed by the Water Pollution Control Directorate, Environmental Protection Service, and approved for publication. Approval does not necessarily signify that the contents reflect the views and policies of the Environmental Protection Service. Mention of trade names or commercial products does not constitute recommendation or endorsement for use.

ABSTRACT

The purpose of this report is to review the new technology and cost considerations of pulping spent liquor recovery. Only in-place technology is considered. The cost information deals with evaporation through incineration and recovery. Special emphasis is placed on the recovery of liquor from high yield pulping.

The report presents an extensive review of proved new technology available to the industry and couples this information with sufficient details of theory to thoroughly explain the significance of these new developments as they are applicable to the Canadian pulp industry. Special emphasis is given to the Marathon Engineering Inc. (MEI) process and other new recovery concepts. Mill energy considerations are also covered.

The cost of liquor recovery is presented for seven different case examples. In these examples the mill production pulping base and yield are considered to be representative of existing or projected sulphite mill conditions in Canada. Cost data are presented for the MEI process, fluidized bed process, and more conventional liquor recovery involving a standard set of evaporators and burning in conventional recovery furnaces.

RÉSUMÉ

Le présent rapport passe en revue les nouvelles techniques et les coûts de récupération de la lessive résiduaire des usines de pâtes. Il traite uniquement des techniques éprouvées et des coûts associés à l'évaporation par incinération et à la récupération. Il insiste sur la récupération de la lessive émanant des usines de pâte à haut rendement.

L'analyse très poussée des techniques s'accompagne de suffisamment de données théoriques pour en établir l'importance pour l'industrie canadienne des pâtes. Il est question, en particulier, du procédé de la Marathon Engineering Inc. (MEI), ainsi que d'autres méthodes nouvelles de récupération; le bilan énergétique de l'usine fait aussi l'objet de considérations.

Le coût de la récupération de la lessive est analysé par l'étude de sept cas. Dans ces exemples, le réactif et le rendement sont considérés représentatifs des conditions actuelles ou prévues des usines canadiennes de pâte au bisulfite. Les données économiques se rapportent aux procédés de la MEI et de la combustion en lits imprégnés de fluides ainsi qu'à la récupération plus habituelle de la lessive à l'aide d'une batterie d'évaporateurs et de fours de récupération.

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1 INTRODUCTION

1.1 Assignment and Scope of Work

During 1972 a report was prepared by EKONO, Inc. and published (Report EPS-3-WP-72-2) by the Environmental Protection Service of the Department of the Environment, Ottawa, Canada, on sulphite pulping spent liquor recovery and effluent secondary treatment. That report covered the then current commercially available and industrially utilized methods for spent sulphite liquor (SSL) in-plant recovery and out-of-plant effluent secondary treatment, with respect to process design, operation, performance and cost.

The Environmental Protection Service, recognizing a need to update this original report, assigned Howard Edde, Inc., Consulting Engineers to undertake this project. The focus of this update review is on new technology and cost consideration of pulping spent liquor recovery. Only in-place technology is considered and the report is limited to pulping liquors.

1.2 Study Ground Rules

This report places special emphasis on the recovery of high yield sodium bisulphite spent pulping liquor, pulp washing, the MEI process, and vapour recompression evaporation. However, other sulphite spent liquor recovery practices and pulping yields will also be considered. The factors affecting cost considerations and trends in optional integrated internal spent liquor recovery systems will also be reviewed.

Throughout this report pulp production figures are given in air dry, screened, unbleached short tons, i.e., adt = 1800 lb absolute dry pulp, if not otherwise designated.

The mention in this report of specific suppliers and trade names, and the exclusion of others, does not in any way imply our recommendations or preference, but is only brought about for the sole purpose of process description and equipment illustration.

2 GENERAL

2.1 Basic Process

The sulphite pulping process is a chemical process that converts wood of many different species into a fibrous pulp. Wood chips enter large vessels called digesters and the chemical pulping process (cooking) takes place under pressure in the presence of heat and acidic solutions (cooking liquor).

At the completion of the cooking cycle the mixture of pulp and residue is discharged from the digesters and the pulp is washed and separated from the residue. The residue is called spent sulphite liquor (SSL). The SSL is then concentrated in multiple effect evaporators and burned in a recovery boiler. For most pulping bases the pulping chemicals can be recovered. The chemicals are reused to make up new cooking liquor. The subject of this report is the disposition of SSL by means of a spent sulphite liquor recovery system.

A major factor in the economics of sulphite pulping is the estimated difference between costs incurred with recovery and non-recovery. The cost impact with recovery is a major area of concern in this report.

2.2 Sulphite Pulping in Canada

The pulp and paper industry is Canada's largest industrial employer, providing about 75,000 factory and office jobs and another 90,000 full time and seasonal jobs in forest operations [1]. Figure 1 indicates the changes in production of groundwood, sulphite and kraft pulp occurring over the past 50 years. The Appendix to this report lists all the sulphite mill operations in Canada by province and describes their yield, pulping base, and daily production. A listing of all mills which have liquor incineration systems is also included along with a brief description of the existing systems. Finally, a list of those sulphite mills which have discontinued operation within the last ten years is included. The list of mills with incineration does not include the projected new 250 t/d corrugating medium plant at Cabano, Quebec. This mill will involve the new Owens-Illinois sulphur-free pulp digestion process and a Copeland Fluidized Bed Recovery System [2]. Table 1

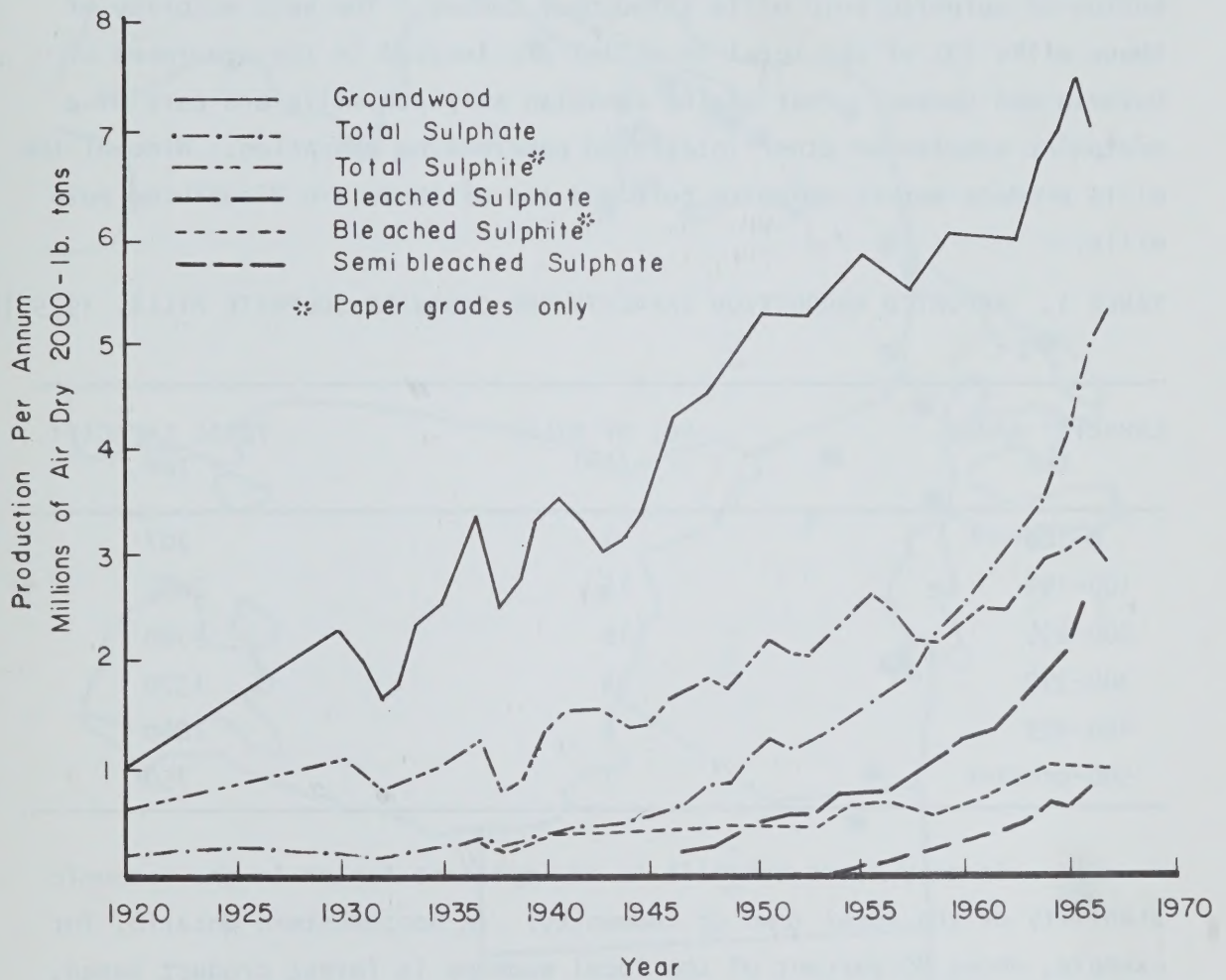


FIGURE 1. PULP PRODUCTION TOTALS FOR CANADA FROM 1920 TO 1967.
(Figures given in millions of air dry 2000 lb tons
per annum)

summarizes the production of Canadian sulphite mills into capacity ranges. A large number of the sulphite mills are of small or medium capacity, with 32 of the 44 mills having a production capacity in the range of 100 to 300 tpd. Figure 2 illustrates geographically the distribution of sulphite pulp mills throughout Canada. The vast majority of these mills (32 of the total 44 mills) are located in the provinces of Ontario and Quebec. Most of the Canadian sulphite mills are part of a newsprint complex or other integrated papermaking operation. Nine of the mills produce market sulphite pulp and two of these are dissolving pulp mills.

TABLE 1. REPORTED PRODUCTION CAPACITY FOR CANADIAN SULPHITE MILLS, 1975 [8]

CAPACITY RANGE tpd	NO. OF MILLS (44)	TOTAL CAPACITY tpd
0-100	3	207
100-199	16	2084
200-299	16	3780
300-399	4	1370
400-499	4	1840
500-greater	1	750

In many cases the mill is an important factor in the economic stability of the local town or community. In northwestern Ontario, for example, about 80 percent of the local economy is forest product based.

2.3 Development Trends in Canada

During the past decade many of the older sulphite mills in Canada have been caught between the need for modernization and effluent control facilities. In fact, six Canadian sulphite mills have discontinued operation during the past decade. Concurrently, during this period industry profits were low and the Canadian economy as a whole was in a recession. Especially for the small to medium sulphite pulp mills, this has resulted in a delay in improvements to their spent liquor collec-

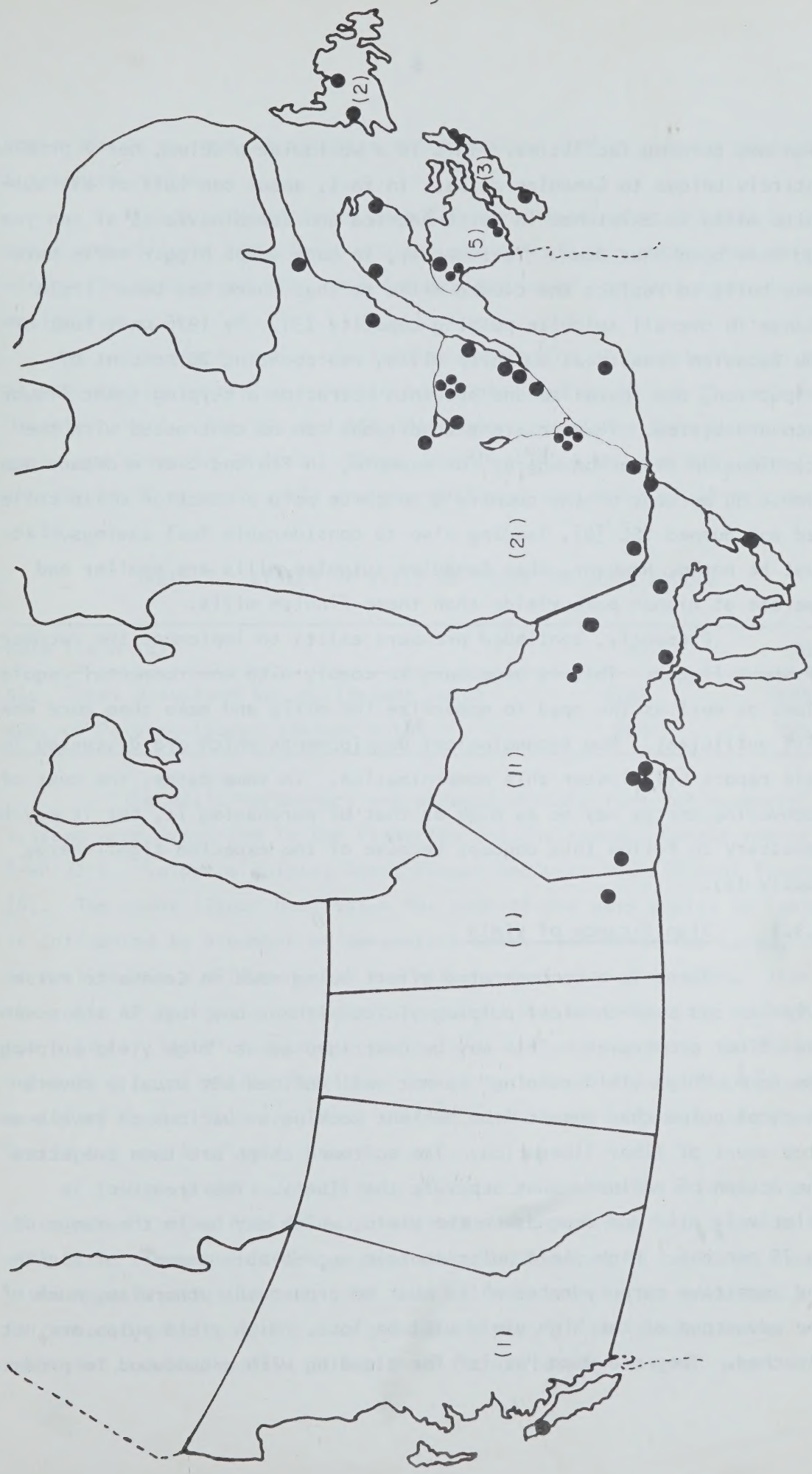


FIGURE 2. GEOGRAPHICAL DISTRIBUTION OF SULPHITE MILLS THROUGHOUT CANADA

tion and burning facilities. This is a worldwide problem, not a problem entirely unique to Canadian mills. In fact, about one half of all sulphite mills in existence in North America and Scandinavia as of ten years ago have been shut down. Fortunately, in many cases bigger mills have been built to replace the closed mills so that there has been little change in overall sulphite pulping capacity [3]. By 1976 only four of the Canadian industrial sulphite mills, representing 20 percent of production, had installed and put into operation a pulping spent liquor recovery system. These current conditions can be contrasted with the Scandinavian situation where, for example, in Finland over a decade ago almost 90 percent of the country's sulphite pulp production mills collected and burned SSL [4], leading also to considerable fuel savings. It must be noted, however, that Canadian sulphite mills are smaller and operate at higher pulp yields than these Finnish mills.

Presently, continued pressure exists to implement the recovery of spent liquor. This is necessary to comply with environmental regulations as well as the need to modernize the mills and make them more energy self sufficient. New technological developments which are discussed in this report will foster this modernization. In some cases, the cost of recovering energy may be as high as that of purchasing it, but it may be necessary to follow this concept because of the expected tight energy supply [5].

2.3.1 Significance of yield

There is a concentrated effort being made in Canada to raise sulphite and semi-chemical pulping yields without any loss in the essential fiber properties. This may be described as "high yield pulping". The term, "high yield pulping" is not well defined but usually covers chemical pulps that result from lenient cooking at various pH levels and stop short of fiber liberation. The softened chips are then subjected to the action of refiners that separate the fibers. The treatment is relatively mild and depends on the yield, which may be in the range of 60 to 75 percent. High yield pulps contain appreciable amounts of lignin and sensitive carbohydrates which must be preserved; otherwise, much of the advantage of the high yield will be lost. High yield pulps are not bleached. They are used "as is" for blending with groundwood to produce

newsprint or as neutral sulphite semi-chemical (NSSC) pulp to produce corrugated medium.

Figure 3 illustrates both (a) the relationship between pulping yield and effluent BOD_5 load, and (b) the calculation method for determining the discharged BOD_5 load for any degree of in-plant spent liquor collection.

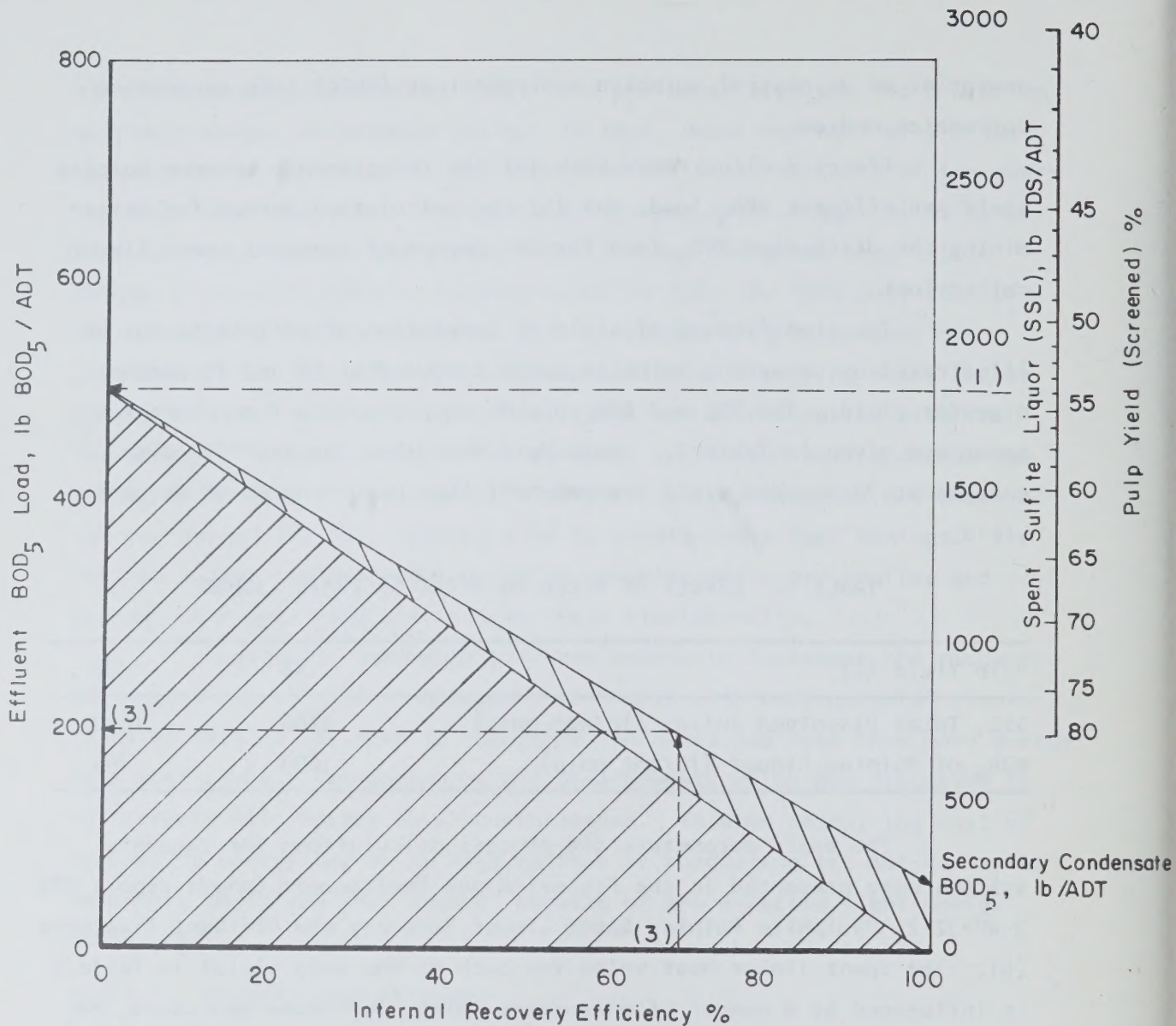
The significance of yield on generation of pollutants can be illustrated by comparing sulphite pulps produced at 48 and 70 percent digester yield. The SSL and BOD_5 discharges resulting from these two cases are given in Table 2. These data show that SSL and BOD_5 discharges at 70 percent yield are one-half that occurring at 48 percent yield.

TABLE 2. EFFECT OF YIELD ON SPENT SULPHITE LIQUOR

Pulp Yield (%)	48	70
SSL, Total Dissolved Solids (lb/adt pulp)	2180	1080
BOD_5 of Pulping Liquor (lb/adt pulp)	600	295

The basic parameters and process calculations for recovery systems were presented in the Fisheries and Environment Canada report EPS 3-WP-72-2, "Sulphite Pulping Spent Liquor Recovery and Effluent Treatment [6]. The spent liquor heat value for each of the pulp yields in Table 2 is influenced by a number of parameters which, for these two cases, may be realized in a difference in the net chemical charge to cook. Thus, the ratio of inorganic to organic compounds may be different, such that for a higher input of make-up chemicals a decrease may occur in the heat value to a point where the liquor cannot be burned without the aid of auxiliary fuel. This influence will be considered in greater detail in Sections 3 and 4 of this report.

High yield sulphite pulp production results in a large fraction of pulpwood in the form of pulp which can be processed into paper at the mill site. Consequently, a greater potential income is realized from a given quantity of pulpwood. The pulp in this case is not marketed. It is consumed on-site along with groundwood pulp in the production of newsprint. Additionally, it has been recognized for several years that



BASIC ASSUMPTION: For the wood raw material and cooking process in question, total BOD_5 of the SSL is a function of pulp yield as indicated.

- EXAMPLE:
- 1) Pulp Yield = 54%
 - 2) Assume Secondary Condensate BOD_5 at 100% Recovery = 60 lb/adt (12% of SSL)
 - 3) Internal Recovery = 70%
 - 4) Read Total Effluent BOD_5 = 490 lb/adt

FIGURE 3. EXAMPLE OF ESTIMATING SSL TOTAL BOD_5 LOAD AS A FUNCTION OF INTERNAL RECOVERY EFFICIENCY

capital investment for recovery and liquor treatment systems is lower for high yield than low yield production. The influence of yield on liquor recovery operating cost differs according to the cooking liquor base. Table 3 presents a comparison of data gathered in 1971 which illustrates these cost differences for several different cooking liquor bases when comparing medium and high yield pulping. Considerable caution is urged in using these unit cost or credit figures since the conversion of a sulphite mill from a polluting one to non-polluting one often entails a change in the base chemical, and it is virtually impossible to generalize the cost of chemical recovery [7]. This subject will be considered in greater detail with current updated information in Section 4.

TABLE 3. SULPHITE PULP WASTE LIQUOR TREATMENT AND CHEMICAL RECOVERY, 1971 CAPITAL AND OPERATING COST RANGES*

Cooking Liquor Base	Pulp Yield %	Fixed Capital Investment		Operating Cost	
		(\$ Million)		(\$ per ton pulp)	
		Pulp Mill Capacity (Unbleached tons/day)		Pulp Mill Capacity (Unbleached tons/day)	
		100	700	100	700
Sodium **	48	1.3	5.0	16.70	8.80
	72	.8	3.0	11.70	4.80
Ammonia	44	3.5	14.9	11.80	0.50
	72	2.4	10.3	12.50	3.00
Magnesium	47	4.3	17.4	8.10	(8.10)***
	60	3.6	14.5	15.00	0.10

* Mid 1971 capital and operating costs

** Na_2SO_4 byproduct, Fluidized Bed Process

*** Credit

Source: Arthur D. Little, Inc., estimates

Recent investigations of high yield pulping have also indicated that in addition to reducing BOD_5 discharges, lower toxicity is realized [8]. Figure 4 shows the potential BOD_5 loads from Canadian sulphite mills which could result from higher pulping yields. Column 1 represents all Canadian sulphite mills presently operating and considers the minimum yield for all mills being increased to 50 percent yield, as well as

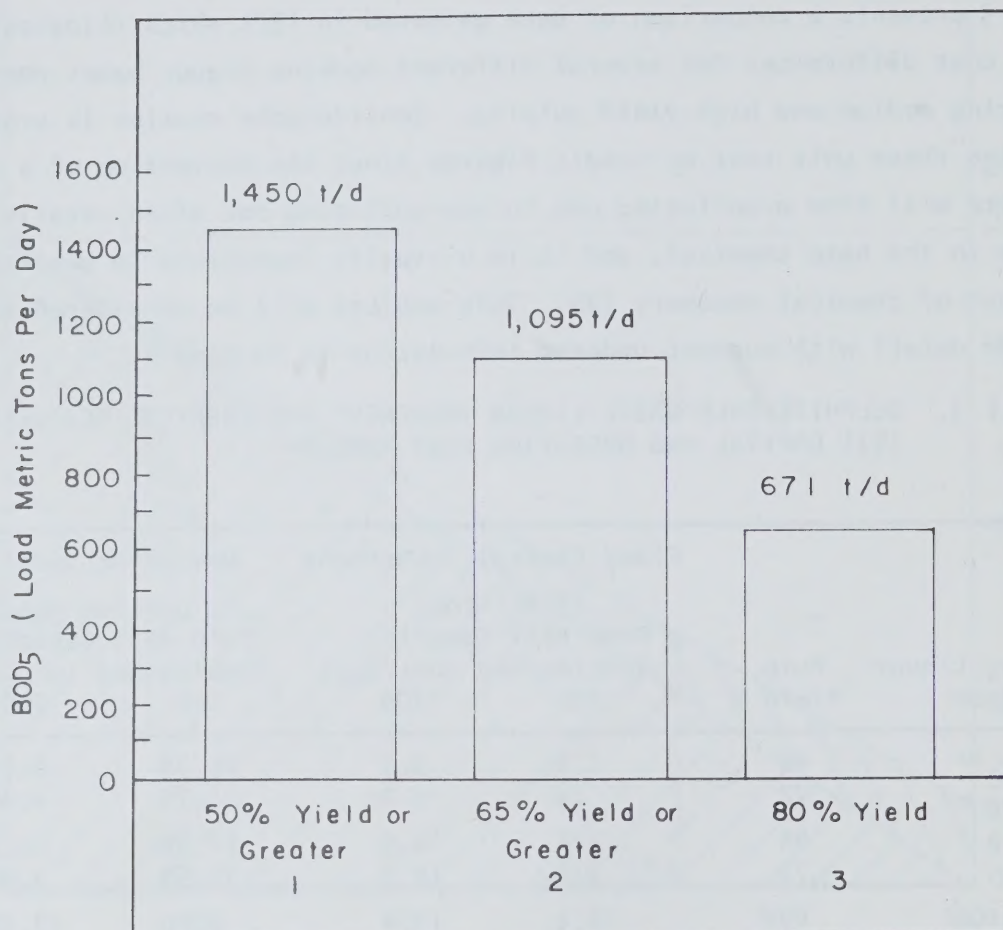


FIGURE 4. POTENTIAL BOD₅ LOADS FROM CANADIAN SULPHITE MILLS BY CONVERSION TO HIGHER PULPING YIELDS

results from mills currently operating at higher yields. Column 2 shows the BOD₅ load if the minimum yield at all mills were increased to 65 percent, again recognizing that some mills currently operate at higher yields.

Despite the favorable effects of high yield operation on toxicity, the reduction in toxicity is not sufficient to meet Fisheries and Environment Canada pulp and paper effluent regulations. Table 4 presents data illustrating the higher the yield, the better the reduction in toxicity. A very large reduction in toxicity occurs at 80 percent yield. However, the effluent is still highly toxic.

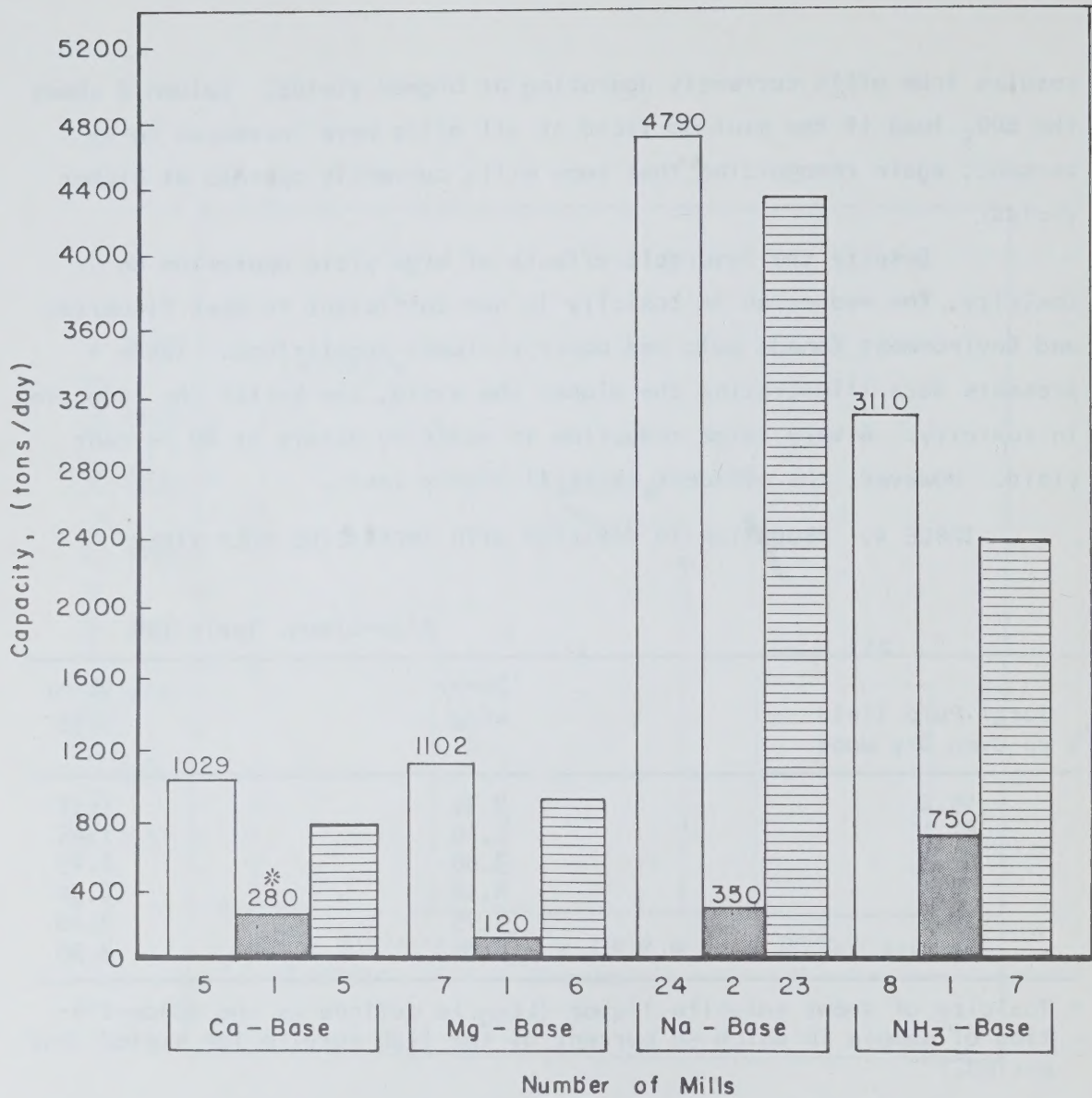
TABLE 4. REDUCTION IN TOXICITY WITH INCREASING PULP YIELD

Total Pulp Yield % on Oven Dry Wood	Blow-Liquor Toxicity*	
	24-hr LC ₅₀ %	96-hr LC ₅₀ %
59.2	2.30	1.55
61.1	2.10	1.45
67.5	3.60	2.45
72.5	4.50	2.90
77.9	4.75	3.00
81.5	5.85	3.90

* Toxicity of spent sulphite liquor (LC₅₀ is defined as the concentration of sample in which 50 percent of the fish survive for stated test period.)

2.3.2 Current status of sulphite pulp manufacturing in Canada

At present, 44 sulphite mills are operating in Canada. This does not include the new 250 t/d Owens-Illinois sulphur free pulping process plant at Cabano, Quebec. The production tonnage and status of SSL treatment at these mills is summarized in Figure 5. The extent of chemical recovery for these mills is described in the Appendix. The production tonnage by pulping base is 10 percent calcium, 11 percent magnesium, 48 percent sodium, and 31 percent ammonium. The number of mills pulping by each base is five mills on calcium base, seven mills on magnesium base, 24 mills on sodium base, and eight mills on ammonia base.



LEGEND

Total



SSL Treatment



No Treatment



* About 1/3 of one mills
SSL recovered to produce by-products

FIGURE 5. CURRENT STATUS OF CANADIAN SULPHITE MILLS

3. NEW TECHNOLOGY

3.1 Pulp Washing and Technology Limits

3.1.1 Application

Pulp washing is the process designed to remove the liquor and other foreign material from the pulp suspension after a completed cook. For economical reasons this must be done without undue dilution of the cooking liquor.

Pulp washing in itself consumes no heat and requires only a minimum amount of power, but it is of key importance to the operation of the recovery system. With an increasing amount of wash water the dry solids losses will decrease and the recovery of heat and chemicals will increase, but at higher evaporation demand.

3.2.1 Technology available

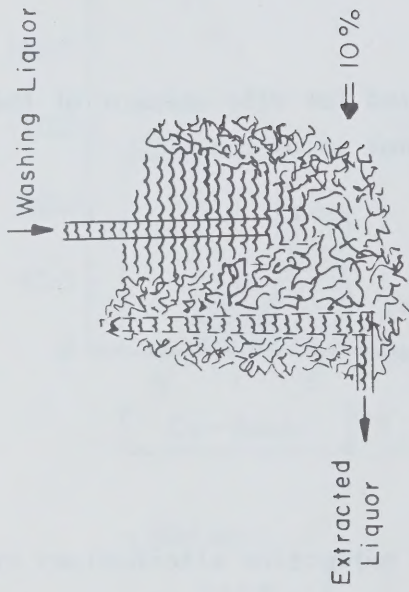
The pulp washing systems normally used for displacement of the spent liquor from sulphite pulp are displacement in or on:

- batch digesters,
- blowpits, one or multi-stage,
- rotary filters, one or multi-stage,
- continuous digesters and diffusers,
- batch or continuous diffusers,
- presses,
- combinations of the above.

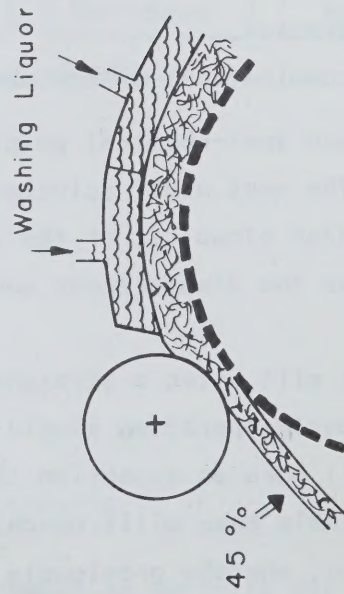
In sulphite and semi-chemical pulping several collection alternatives are available. The most usual solutions are batchwise liquor collection in digesters and/or blowpits, or the use of rotary filters and presses. Figure 6 shows the displacement conditions which occur in some of these systems [9].

The mill using a straightforward approach of liquor collection on a continuously operating simple washing line (say rotary filters) is, however, still more an exception than a rule. This is due to the fact that most soluble base mills operating today are converted from calcium base operation, whereby previously used liquor collection equipment is incorporated in the washing system in order to save capital costs. The

RADIAL WASHER



WASHING PRESS



WASH FILTER

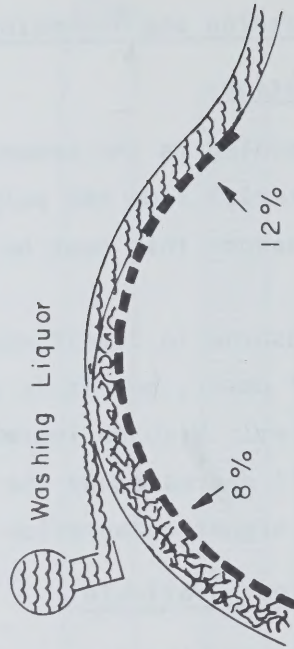


FIGURE 6. DISPLACEMENT CONDITIONS ON VARIOUS PULP WASHING EQUIPMENT

main features of the different liquor collection systems are discussed in the following [10]:

Batch digesters - Displacement of the spent liquor directly in the batch digester is used widely in European sulphite mills. The digester is used for one or more displacement stages. The washing is further completed in blowpits or on rotary filters. This practice requires more digester capacity in terms of digester volume per ton of pulp produced.

Although digester displacement might be of limited interest in this connection, it is worth noticing that in efficiency it easily replaces one rotary washer.

Blow pits and diffusers - Blow pits exist in many different shapes and materials of construction. Commonly, the blow pit is a round or rectangular, wooden or wood lined tank with a perforated plate slightly above the bottom to facilitate pulp washing. As such, blow pits have been used for displacement of the spent cooking liquor since the early days of sulphite pulping.

Blow pit washing can well be combined with digester displacement, as previously mentioned, or with one or more additional filter washers. Some difficulties can be experienced in the transition from batch to continuous operation, but modern instrumentation and remote control will minimize the problem.

A principal flow diagram for spent liquor collection in blow pits is shown in Figure 7. This system applies two stages of washing for each blow, and the intermediate wash liquor tank is baffled to retain the concentration gradient. With a well designed and operated system of this kind, liquor collection efficiencies in the order of 85 percent can be reached, without excessive dilution.

Since blow pit washing systems almost without exception have been developed from existing facilities, maintenance cost generally is rather high. Concrete does not stand up to hot concentrated spent sulphite liquor, and wood deteriorates rather fast. Stainless steel has been used successfully, especially for the sieve plate, but at high cost.

Rotary washers - Figure 8 shows a principal flow diagram for spent liquor collection in a rotary washer plant, featuring, in this example, three two-zone filters in series. The pulp and wash liquors follow a counter-

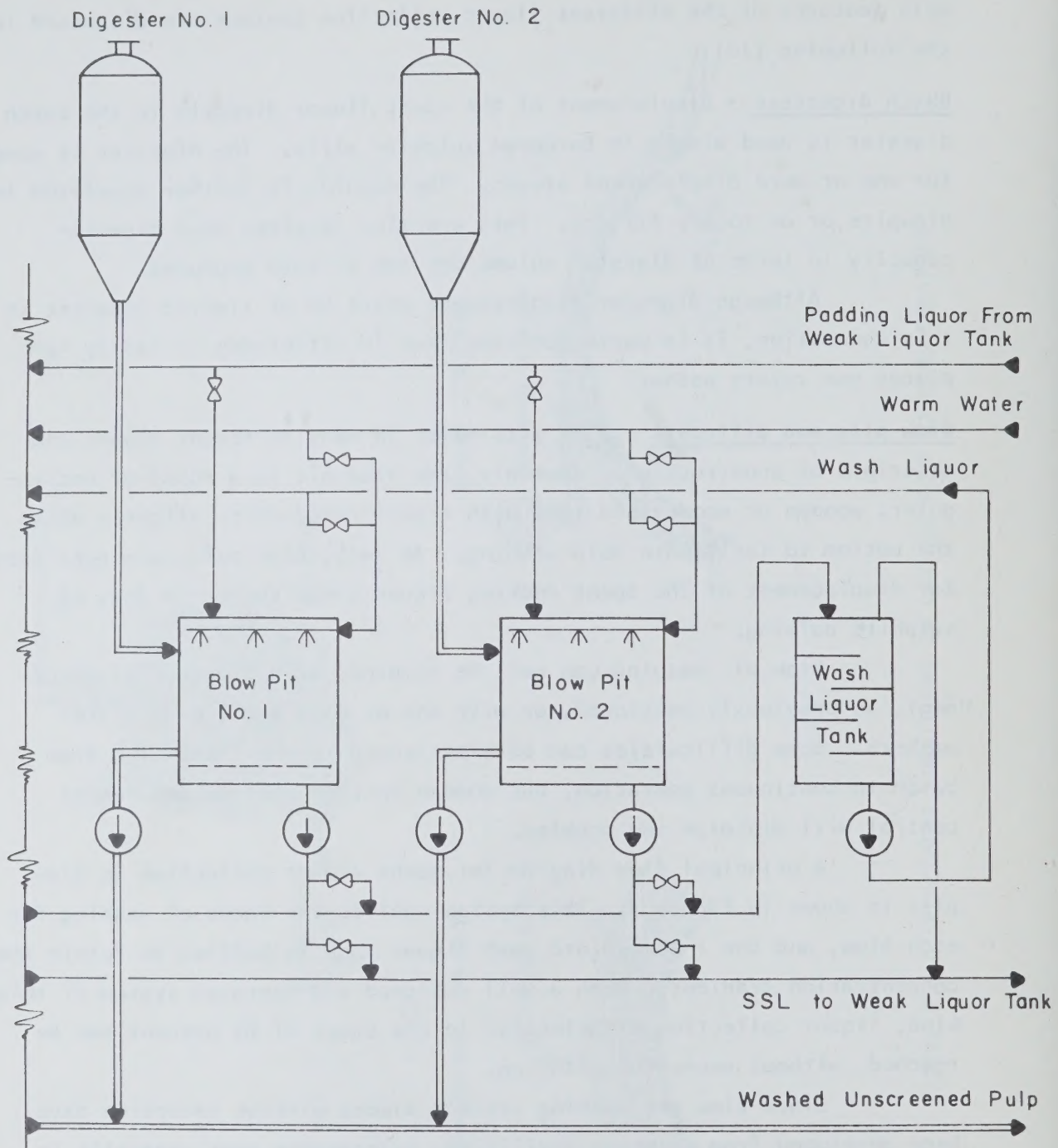
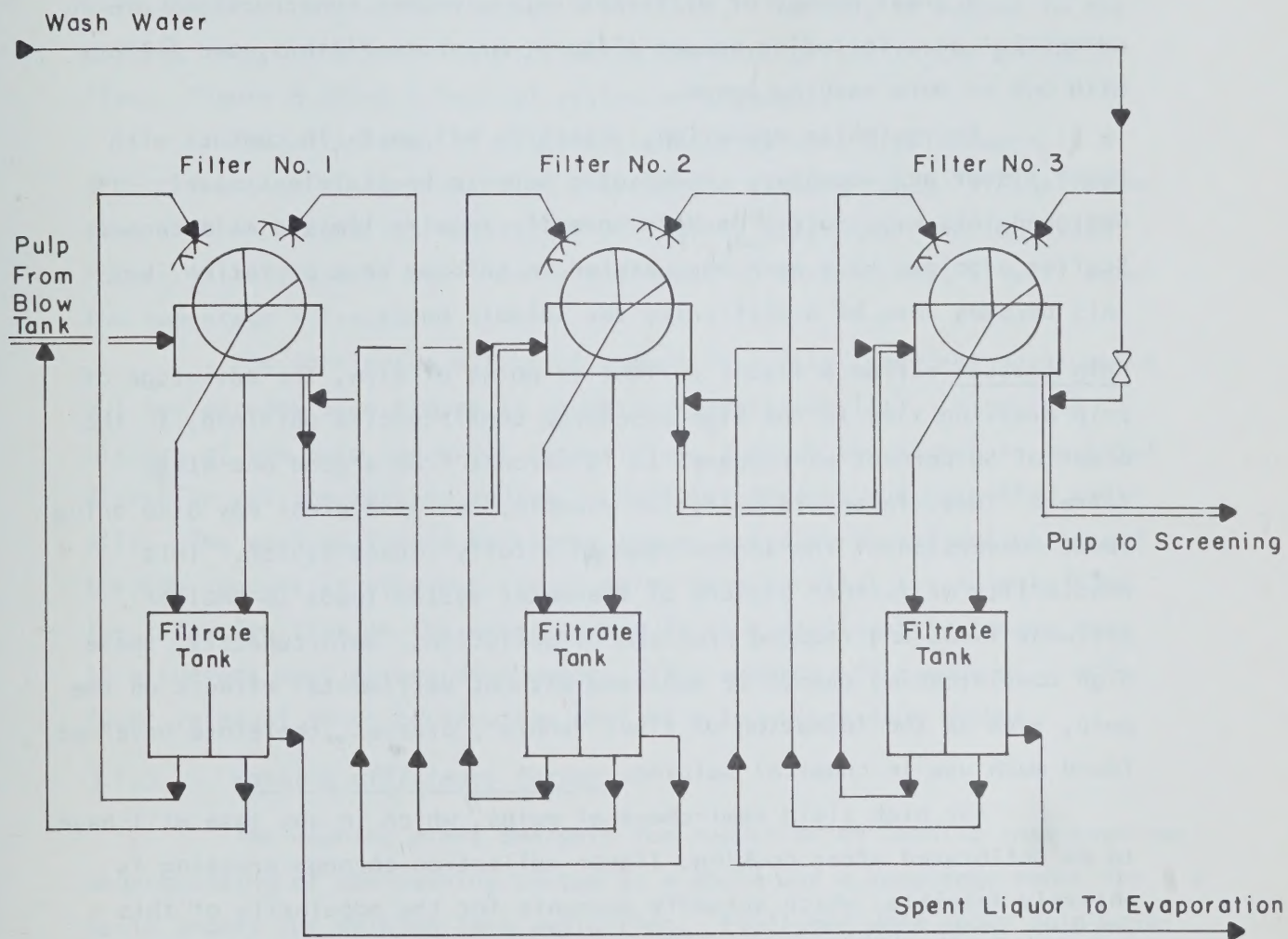


FIGURE 7. FLOW DIAGRAM FOR SPENT SULPHITE LIQUOR COLLECTION IN BLOW PITS



Plant Type: Three Rotary 2-Zone Filters

FIGURE 8. FLOW DIAGRAM FOR SPENT SULPHITE LIQUOR COLLECTION IN A ROTARY WASHER PLANT

current pattern. The efficiency of a plant of this type will, as usual, vary with equipment performance and accepted liquor dilution. The use of three filters is generally economically justified at collection efficiencies approaching or exceeding 90 percent.

A great number of different rotary washer constructions are in commercial use, including vacuum filters, pressure filters, and filters with one or more washing zones.

For sulphite operation, generally all parts in contact with spent liquor and secondary condensates have to be stainless steel. If designed this way, rotary washers normally require limited maintenance. Scaling problems have been encountered in calcium base operation, but this becomes less of a difficulty for soluble bases.

Pulp presses - From a liquor collection point of view, the advantage of pulp pressing lies in the high discharge consistencies obtained, in the order of 50 percent as compared to 15 percent from a good one-stage filter. Thus, in an old mill, for example, the wash press may also bring about conversion of the screen room to a fully closed system. This possibility of further closure of the water system leads to smaller effluent flows and reduced problems of pollution. Unfortunately, these high consistencies cannot be achieved without detrimental effects on the pulp, such as the formation of fiber "knots"; presses, therefore have not found much use in chemical pulping.

For high yield semi-chemical pulps, which in any case will have to be defibrated after cooking, liquor collection through pressing is entirely feasible, which actually accounts for the popularity of this approach in the NSSC process. In Canada, pulp presses would be very attractive for a high yield process as well because high consistency refining develops good strength properties and reduces refiner energy requirements. High yield sulphite mills without liquor recovery quite often have pulp presses in Canada.

Pulp presses are expensive pieces of equipment, and the power consumption, as well as the cost of maintenance, is high. Thus, these costs have to be balanced against possible advantages in evaluating the feasibility of using this equipment for liquor collection.

Displacement in continuous digesters and diffusion washers - Counter-current displacement in continuous digesters has proven a very efficient

washing method in sulphite pulping. Displacement of the liquor is made in a washing zone located immediately after the cooking zone in the digester. The pulp and liquor from the cooking stage are cooled using quenching circulation, and the mixture of the original liquor and the wash liquor is extracted through a strainer. The pulp is washed in the washing zone using wash liquor flowing counter-current to the product flow. Figure 9 shows a typical system arrangement.

The efficiency of digester counter-current displacement is a direct function of the retention time in the washing zone. If the retention time is around 1.5 hours, the efficiency will equal one two-stage rotary filter, and if the time is increased to three hours, it will equal two two-stage filters.

The continuous diffusion washer is a relatively new development but has already been proven in commercial operation [11]. Here, the mixture of the pulp and spent liquor flows through the washer in an axial direction and the washing medium is diffused through the main flow radially. The washing liquid and spent liquor are distributed and collected through concentric strainer rings, which move parallel to the pulp flow. The retention time in the washing zone is thus considerably longer than in a conventional rotary drum washer. The washing efficiency of a diffuser is equal to or better than that of a two-stage drum washer.

3.1.3 Washing efficiency factor

The washing plant analysis for reduction of washing loss requires understanding of the washing system as a whole and a knowledge about the basic theory for washing loss evaluation. Published work about pulp washing gives detailed theory for determination of washing efficiency and washing losses [12].

The most important parameters influencing washing loss are the equipment efficiency and dilution factor. Figure 10 shows the washing loss for different equipment efficiencies and dilution factors (expressed as excess wash water as defined below).

The Norden efficiency factor, E , is often used as a measure of the washing equipment efficiency. The efficiency factor, E , of a washing stage, or an entire washing plant, is defined as the number of counter-current ideal mixing stages (where the departing wash liquor has the same

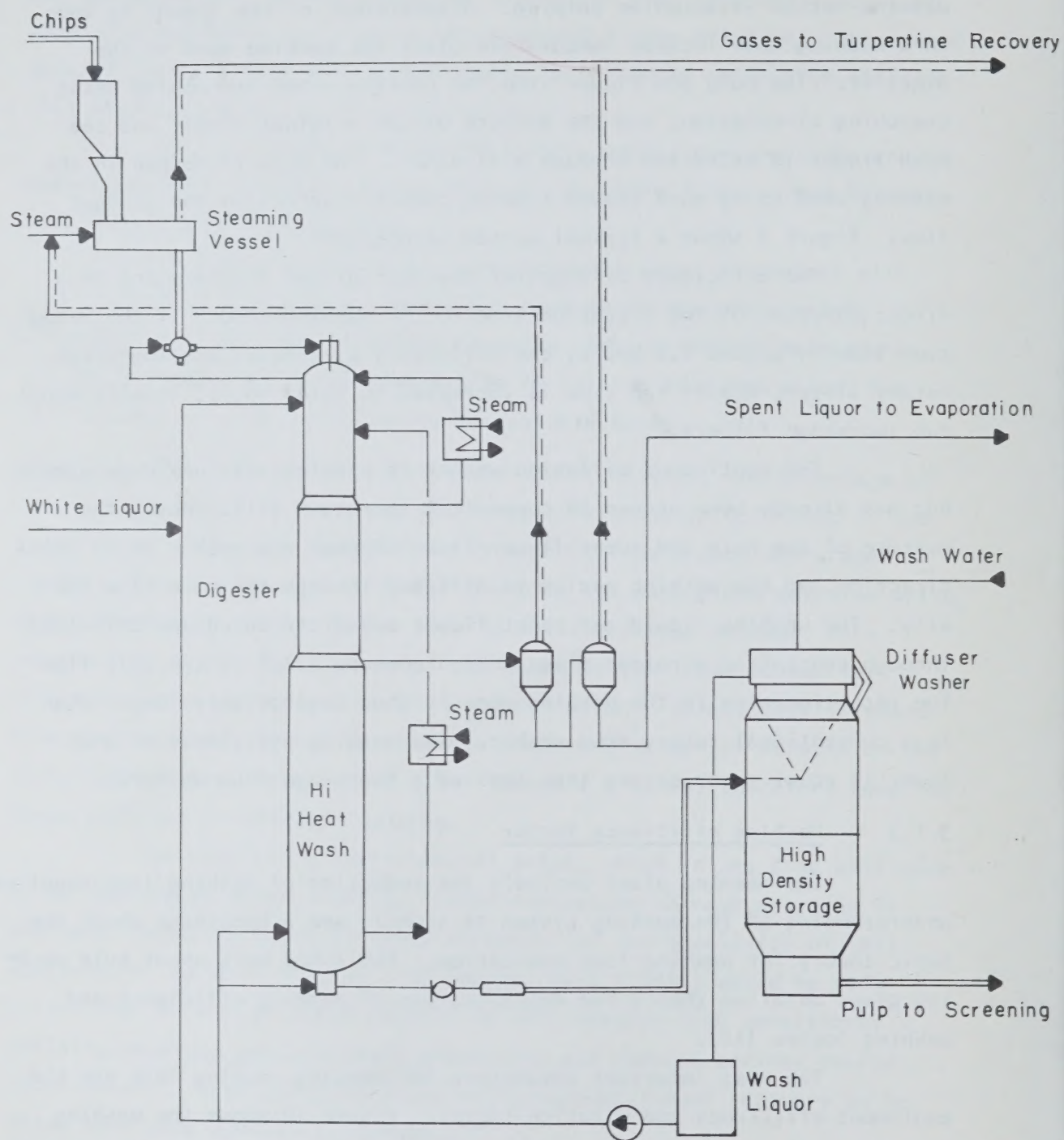


FIGURE 9. FLOW DIAGRAM FOR SPENT SULPHITE LIQUOR COLLECTION IN HIGH HEAT AND DIFFUSER WASHER

concentration as the liquor in the discharged pulp) with the same discharge consistency as the real stage, which are required to give the same performance as the stage or plant.

The efficiency factor concept has several important features which include:

- (a) For most types of washing equipment, E is only slightly dependent on the dilution factor, which is one of the most important design and operating variables.
- (b) E is proportional to plant efficiency and cost capital.
- (c) Use of E simplifies the comparison of different types of equipment.
- (d) It simplifies calculations of washing plant performance.

E values for different types of liquor collection equipment are given in Table 5.

TABLE 5. EFFICIENCY FACTORS FOR DIFFERENT TYPES OF WASHING EQUIPMENT

Equipment	Discharge Consistency %	Typical E-Values of Stage
Batch Digester (1 stage)	10	3.5
Blow Pit (1 stage)	10	3.5
1-stage Rotary Washer	12	2.5
	15	2.5
2-stage Rotary Washer	12	3.5
	15	5.0
Press	35	1.0
	50	1.0
	50	0.5
Digester Hi-Heat Wash		
- Retention time 1.5 hr	10	5.9
- Retention time 3 hr	10	9.6
Continuous Diffuser	9	5.0

The dilution factor, EW (expressed as excess wash water amount), is determined as follows:

$$EW = \text{wash amount} - \text{amount of liquor out with pulp.}$$

This means that EW is a function of not only the shower flows, but is very heavily dependent on the discharge consistency. Usually, the washing plant is operated with constant wash water volume per production unit. A change in discharge consistency will then cause a change in the dilution factor. For example, assume the wash water amount is 10 t water/tp (tons of water per ton of pulp). The dilution factor changes from 1 m³/tp to 4.3 m³/tp when the discharge consistency changes from 10 to 15 percent. Consequently, the wash water amount as well as discharge consistency should be checked, in order to operate the washing plant at the correct dilution.

The relationship between washing yield, efficiency factor and excess wash water amount is defined from the equation given and shown graphically in Figure 10.

Although pulp washing normally can be regarded as a rather closed unit operation, the main effluent sources being occasional spills and gland leaks, this is not always the case for sulphite pulp washing. Because of the rather low liquor collection degree used, the amount of spent liquor remaining in the pulp is rather high, even after complete liquor recovery, and will lead to excessive foaming problems in the screen room and an increased bleaching chemicals consumption for bleached grades.

3.1.4 Calculations of washing plant performance

There are several methods for determining the efficiency of the washing system. The method presented here is based on the Norden efficiency factor theory of leaching which was defined previously in section 3.1.3. It does not rest on any assumptions regarding operating principle or the construction of the equipment. It is rather a mathematical theory to simplify the calculations of multi-stage counter-current systems using a knowledge of the performance of single stages in the system. It has, however, great merits as an efficient method for condensing performance data and correlating them with relevant performance parameters [12].

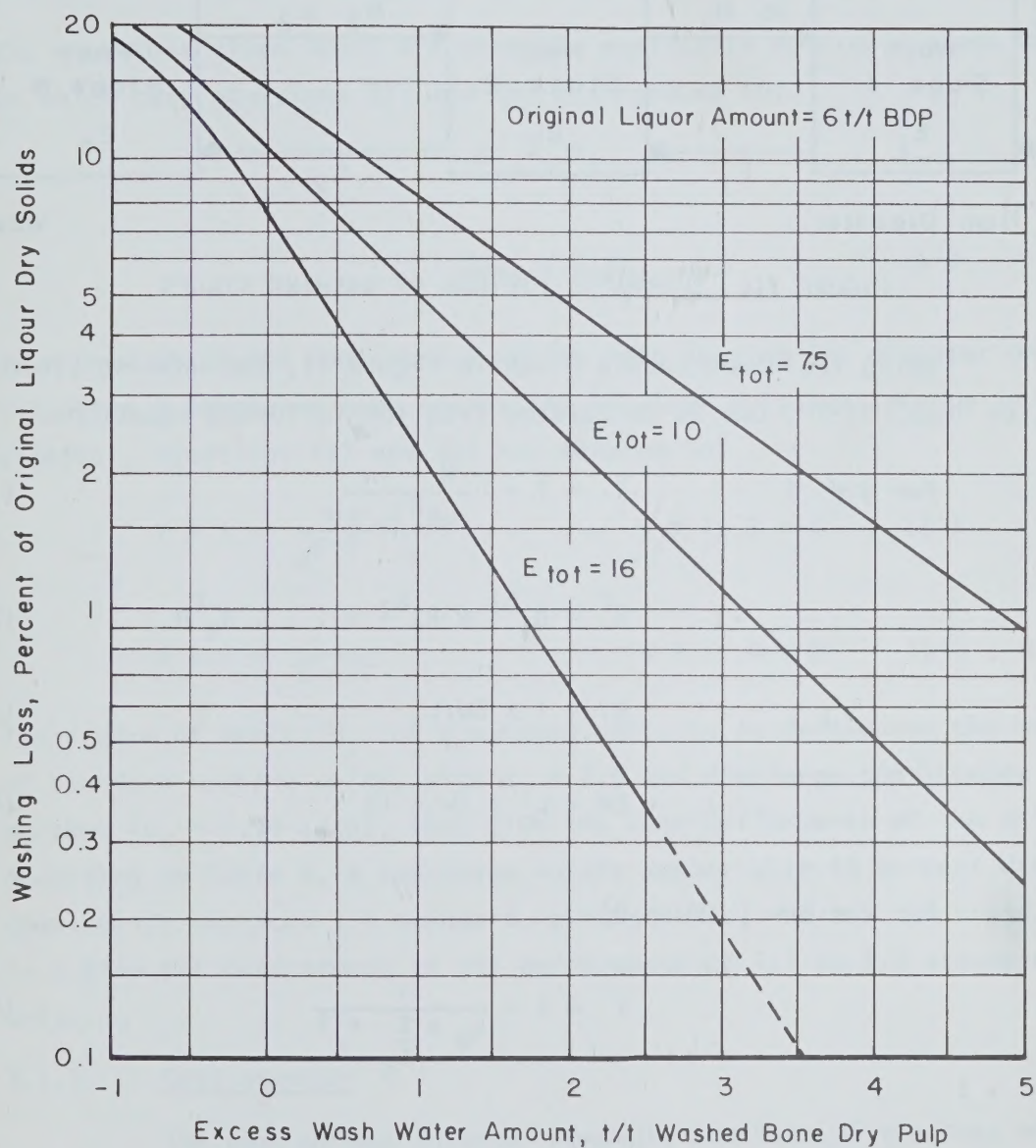


FIGURE 10. WASHING LOSS FOR DIFFERENT WASHING EQUIPMENT EFFICIENCIES AND DILUTION FACTORS

Schematically, the washing plant is shown in Figure 11.

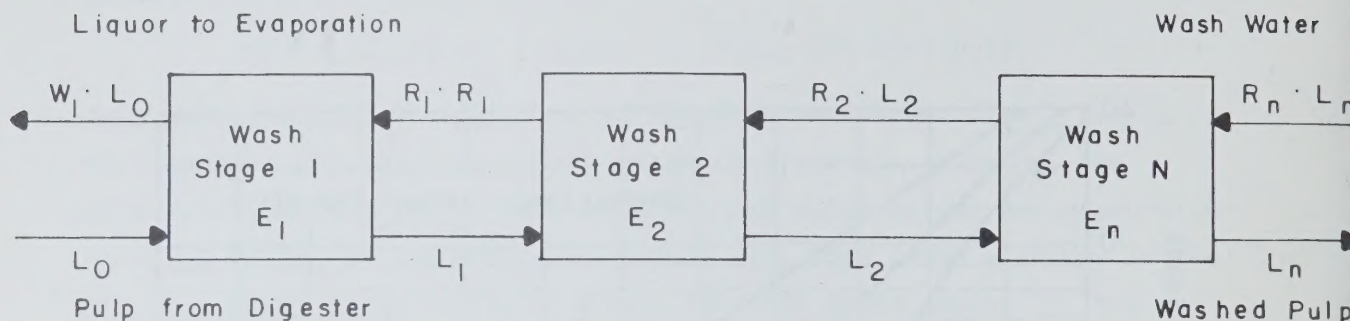


FIGURE 11. SCHEMATIC PICTURE OF WASHING STAGES

Using the nomenclature given in Figure 11, the washing yield, "Y", or loss, "1-Y", can be calculated from the following equations:

$$\text{For } W \neq 1 \quad Y = 1 - \frac{W - 1}{WR^E - 1} \quad (1)$$

$$R^E = R_1^{E_1} \times R_2^{E_2} \times \dots \times R_N^{E_N} \quad (2)$$

$$R_i = 1 + EW/L_i \quad (3)$$

$$EW = L_0 \times (W - 1) \quad (4)$$

For $W = R = 1$ ($EW = 0$)

$$Y = 1 - \frac{1}{L_0 \times \frac{E}{L_N} + 1} \quad (5)$$

$$E/L_N = E_1/L_1 + E_2/L_2 + \dots + E_N/L_N \quad (6)$$

where: E = Norden washing efficiency factor,
 EW = excess wash liquor flow,
 L_i = flow of liquor with pulp from stage i ,
 L_0 = amount of original liquor to washing plant,
 R = wash liquor ratio = wash liquor to stage or plant/liquor with pulp leaving the stage or plant,

W = liquor weight ratio = amount of recovered liquor / original liquor amount,

Y = washing yield = recovered dry solids or chemicals amount / original dry solids or chemicals amount.

The equations given above are directly applicable to single stage washing, in which case equations (2) and (6) are reduced to:

$$R^E = R_1^{E_1} \quad (2')$$

$$E/L_N = E_1/L_1 \quad (6')$$

In a case where the consistency of the pulp leaving the digester or blow pit after a completed wash equals the original pulp consistency in the digester, equations (1) and (5) are reduced to:

$$Y = 1 - \frac{W - 1}{W^{E+1} - 1} \quad W \neq 1, R = W \quad (1')$$

$$Y = 1 - \frac{1}{E + 1} \quad W = 1, R = W \quad (5')$$

The figure of merit, M_i , of a washing unit, i , is defined as the number of standard washing units, with $E_i = 2.5$ and discharge consistency 12 percent ($L_i = 7.33$ t/tp), that give the same performance as the unit, i . According to Table 6, a two-stage rotary washer with 12 percent discharge consistency replaces 1.4 standard washing units, whereas one-stage washing in a blow pit corresponds to the performance of 1.1 to 1.2 standard washing units.

3.1.5 Cost aspects

The cost of the internal recovery consists of the cost of the main equipment itself, but the effect of the recovery measures on the total system must also be considered. For example, higher spent liquor collection efficiency achieved by changing the equipment and operation in the washing plant will affect the system throughout the whole recovery cycle. In the case of a new mill to be built, the additional capacity is available at the moderate marginal cost of the equipment. In the case of an existing mill, the problem is more complicated because of the limited capacity of the existing equipment, and the high initial price of the additional capacity.

TABLE 6. E-VALUES FOR DIFFERENT TYPES OF LIQUOR COLLECTION EQUIPMENT

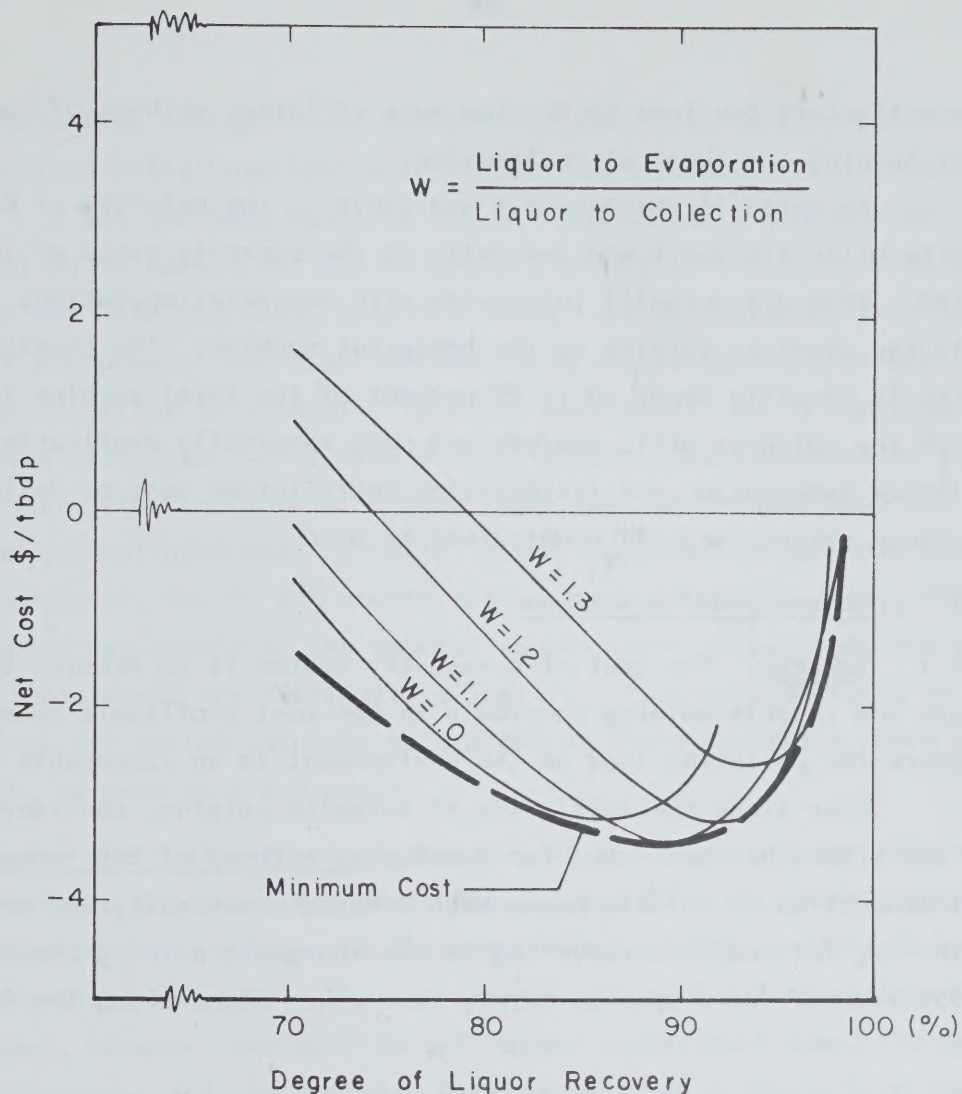
Equipment	Discharge Consistency %	Typical E-Values E_i	Figures of Merit, M	
			EW=0	EW=2
Batch Digester (1-stage)	10	2.5-3.5	0.81-1.14	0.83-1.16
Blow Pit (1-stage)	10	3-4	0.98-1.30	1.00-1.33
1-stage Rotary Washer	12	2.5	1.0	1.00
	15	2.5	1.29	1.25
2-stage Rotary Washer	12	3.5	1.4	1.4
	12	5.0	2.0	2.0
Press	35	1.0	1.58	1.22
	50	1.0	2.93	1.84
	50	0.5	1.47	0.92
Digester Hi-Heat Wash				
- Retention time 1.5 hr	10	5-6	1.63-1.95	1.66-2.00
- Retention time 3.0 hr	10	8-10	2.61-3.62	2.66-2.33
Continuous Diffuser	9	5	1.45	1.52

The total economic system affected by changes in the washing plant is thus rather complex. For practical work, however, simply summing up the different cost items as a function of two main variables, dilution and washing efficiency, is very instructive.

For a given set of washing equipment and evaporation capacity, there is always an economic optimum point of operation determined by the price of steam and lost chemicals. This is illustrated in simple form in Figure 12, which shows the plotted results of a wash filter and four effect evaporators with $W > 1.0$, 1.1, 1.2, and 1.3 (W explained in Figure 12). The data show that the minimum SSL recovery cost occurs at different degrees of recovery for the W values.

3.2 Review of Evaporation and Incineration Practices

The evaporation and incineration practices used in the sulphite and semi-chemical pulping industry differ widely from process to process



PROCESS	Mg. BISULPHITE (NO EXISTING RECOVERY)
PRODUCTION	10 T BDP/HR
RAY MATERIAL	SOFTWOOD
COOKING YIELD	53%
SCREENED YIELD	52%
BLEACHING LOSS	8-10%
MgO (CHEMICALS)	100 \$/T
S (CHEMICALS)	30 \$/T
FUEL	\$3.5/G CAL (88 /MILLION BTU)
POWER	0.6 /KWH
LABOR (NO RECOVERY)	12.25 \$/T BDP
LABOR (WITH RECOVERY)	13.68 \$/T BDP

T BDP = TON OF BONE DRY PULP

FIGURE 12. NET COST OF LIQUOR RECOVERY

as investigators continue to develop more efficient methods of spent liquor burning and chemical reclamation.

As noted in the Appendix and Table 1, the majority of Canadian sulphite mills are small and generally in the capacity range of 100 to 300 tpd. They are normally integrated with newsprint operations and supply the chemical furnish to the newsprint machine. The chemical furnish is normally about 20 to 25 percent of the total machine furnish. Many of the sulphite mills are old and only marginally profitable. Thus, any liquor evaporation and incineration installation must be designed with these unique facility constraints in mind.

3.2.1 Conventional practices

3.2.1.1 General. The goal of a recovery system is to recover by-product and recycle pulping chemicals in the most profitable manner, and to reduce the polluting load on the environment to an acceptable level.

Ever since the initiation of sulphite pulping, considerable time and effort has been used for developing methods of SSL recovery. The introduction of soluble bases both created a necessity and provided a possibility for economic recycling of the inorganic pulping chemicals. The SSL recovery developments mainly have taken place along the following lines:

- (a) recovery of saleable, basically organic byproducts before treatment of the residue, if necessary, by other methods;
- (b) separation of inorganic cooking chemicals for recycling to the process, before treatment by other methods;
- (c) partial or complete oxidation of all oxygen consuming compounds before discharge of the effluents to receiving waters, with or without recovery of process chemicals;
- (d) thermal oxidation, i.e., incineration, of the SSL dry solids, to separate the inorganic process chemicals for possible recycling, and to convert the organic wood residues to carbon dioxide and water vapor under release of heat.

Byproduct recovery - Many examples can be listed for the use of SSL, or byproducts obtained according to point (a) above. These markets gener-

ally have been insignificant compared to pulp consumption, and only a few specialized mills have been able to continue operation on this basis.

SSL has been used, as is, as road binder and dispersant, and several mills have dried SSL to powder, mainly to reduce cost of transportation. Tanning aids, vanillin, etc., can be extracted from the lignin compounds. SSL also contains fermentable monosaccharides, which can be converted to ethanol. This has been practiced in numerous European mills, but it is interesting to note that the value of the heat obtainable from the sugars in some cases has been higher than the value of the ethanol produced. Finally, other low molecular organic compounds, like methanol, formic and acetic acids, cymen, furfural, etc., can be extracted but generally the capital expenditures to purchase the additional equipment are difficult to justify. Further processing of the residual dry solids can only be avoided in those few cases where all of the SSL is used as a byproduct.

Inorganic chemical recovery - Recovery of inorganic process chemicals according to (b) above has been demonstrated in pilot plant scale. Examples of this include the Abiperm, Pritchard-ORF, and Pritchard-Fraxon ion exchange processes. These processes all suffer from serious disadvantages, however, and only in one rather exceptional case, the Ontario Paper Company, Ltd. in Thorold, Ontario, has ion exchange been successfully used as part of a complex recovery system. Specifically, ion exchange processes only recover the pulping base, whereas recovery of pulping sulphur and heat would require subsequent incineration and regeneration operations [13].

Many other attempts have been made to apply sophisticated chemical or physical methods to the recovery of spent liquor and other process effluents. One of these is the work carried out on pilot plant scale by the former Sulfite Pulp Manufacturers Research League (now part of the Institute of Paper Chemistry) on electrodialysis [14]. High power consumption and difficulties in obtaining byproducts with enough market appeal have prevented the process from obtaining commercial acceptance.

Today, sulphite recovery systems are installed almost solely as compulsory measures for pollution control. In general, they are not justified from an economic point of view alone. The search for less costly alternatives has been left to systems aiming at simple oxidation

of the organic matter only, as defined in (c). Zimmermann has suggested such a process based on 'wet combustion', and a couple of commercial installations were operated for some time [15]. However, these processes do require elevated temperature and pressure conditions, and the corresponding increase in capital cost and operating expenses, in conjunction with operational difficulties and lack of chemicals recovered, outdate the systems.

In summary, it can safely be concluded that of all the processes developed to date for SSL recovery, only liquor combustion in conjunction with heat or chemical recovery, or both, as defined in (d) above has been technically and economically competitive. With very few exceptions, under special conditions, all operating recovery systems the world over are of this type. Thus, the subsequent section on cost in this report deals only with liquor combustion systems.

3.2.1.2 Evaporation of spent liquor. The water content of the spent liquor collected in the pulp washing systems is normally very high and the liquor cannot be used as such as a fuel or incinerated without supplementary fuels. The system used for incineration and the minimum temperature needed in the reaction chamber (furnace) for the combustion will determine the required dry solids content of the liquor. This dry solids content can be achieved using indirect or direct contact evaporation or a combination of both, depending on technical and economic aspects.

For calcium and magnesium liquors, the upper concentration limit is in the order of 55 to 58 percent dry solids from multiple effect or vapor recompression plants. Higher figures have been reported for magnesium liquors when evaporated in a direct contact flue gas scrubber for direct feed to the furnace. Ammonia base liquors frequently have been reported to polymerize at high concentrations and temperatures, severely limiting either plant capacity or maximum attainable concentration. The tendency to polymerize varies from case to case, but mostly it should be possible to operate well-designed ammonia base evaporation plants at plus 50 percent concentrations, possibly as high as 55 percent concentrations.

Sodium base liquors can be concentrated well above 60 percent in indirect contact-type evaporators of conventional design. Higher

values up to 65 percent dry solids can be achieved using concentrator units with efficient forced circulation and washing systems.

The most severe problem in indirect spent liquor evaporation is the formation of scale on the heat transfer surfaces and in the thick liquor piping. The nature of the scaling differs, depending on cooking base used. In sulphite operations the scaling is mainly sulphate precipitated from the liquor. It can be stated rather simply that the main reason for scaling is the inverted solubility characteristics of calcium sulphate with increasing temperature. Irrespective of the pulping base, sulphite liquor always contains calcium and sulphate ions, and the precipitation of CaSO_4 has always been found to occur at some stage of the evaporation. Scaling can be minimized to some extent by pH control of the liquor entering the evaporators. The scaling problem has been handled by several authors in the literature [16, 17].

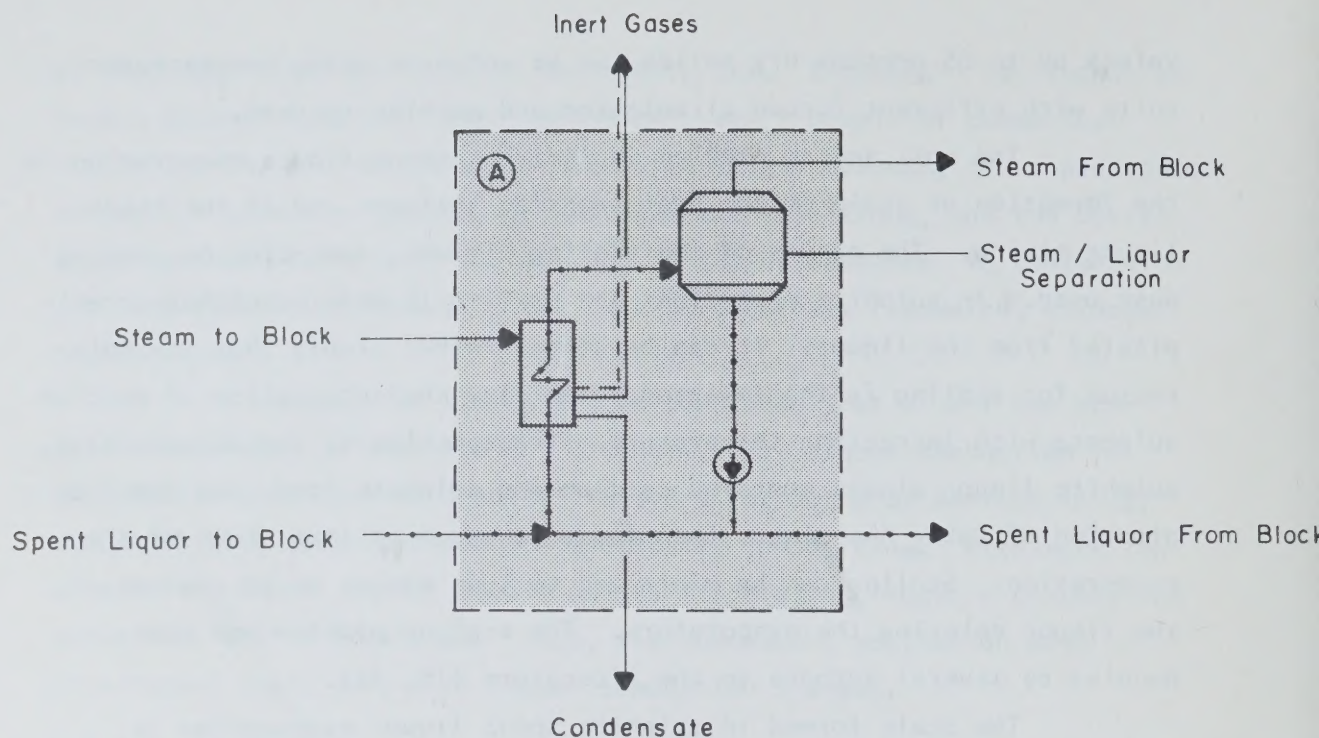
The scale formed in sulphite spent liquor evaporation is soluble in acidic condensates from the plant or in nitric acid. The normal procedure is to wash one evaporation effect at a time with condensates during operation.

3.2.1.3 Equipment. The concentration of spent sulphite liquor is usually performed in multi-stage units, using indirect steam as a heat source. Each stage consists basically of a flash tank from which the liquor is circulated through a heat exchanger, where it is heated by steam. This is illustrated in Figure 13 together with a schematic presentation of a direct contact (flue gas spent liquor) evaporation unit.

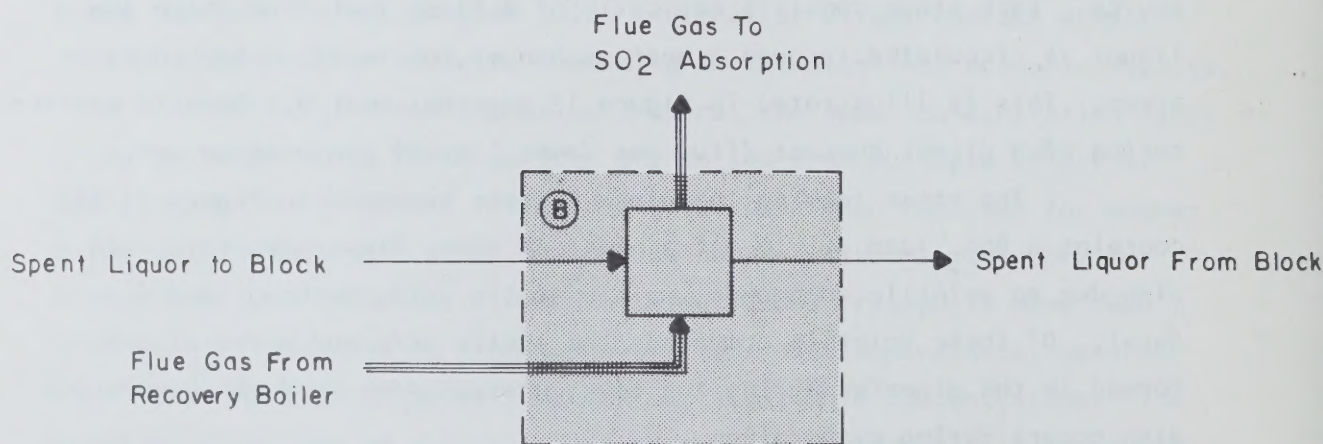
The steam leaving the block balance boundary in Figure 13 (A) contains a BOD_5 load due to entrainment of spent liquor droplets, and also due to volatile compounds, mainly acetic acid, methanol and furfural. Of these volatile compounds the acetic acid and methanol are formed in the digester during the cook, whereas some furfural generation also occurs during evaporation.

Because of the corrosive nature of sulphite liquor, it is impractical to use anything less resistant than stainless steel. Therefore, the capital cost of an evaporator is much higher than for the kraft process. Thus, the number of effects has been reduced to a minimum.

In the earlier developed types of sulphite liquor evaporators, the feed and product pumps and general arrangement of circulating pumps



EVAPORATION USING INDIRECT HEATING LIQUOR



DIRECT CONTACT EVAPORATION WITH RECOVERY BOILER FLUE GASES

FIGURE 13. PRINCIPLE OF SPENT LIQUOR EVAPORATION

and heat exchangers were similar to conventional multiple-effect evaporators used in the kraft mill, but because of the serious scaling problems, special and unique arrangements of evaporators and washing systems were devised.

3.2.1.4 Combustion of spent liquor. Incineration of the spent liquor dry solids today is the most attractive method of disposal. The main objectives of the combustion process are:

- incineration of the organic material in the spent liquor,
- recovery of heat, and
- recovery of chemicals.

The technique of burning sulphite liquor is entirely different from that used for burning kraft liquor. Kraft liquor is sprayed onto the heating surface and is then consumed in a reducing environment to promote good chemical recovery, with secondary combustion air being added over fire to complete combustion of the evolved gases. With sulphite liquor a very fine atomization and spray drying of the fuel is required so that no unburned particles or droplets of liquor reach the heating surface, the liquor being burned completely in suspension.

The technical solution of the burning process chosen depends on the type of cooking process, the desired end product of the combustion and the value of the recovered heat and chemicals. The different types of combustion processes are:

- pyrolysis,
- wet combustion,
- combustion in reductive conditions, and,
- combustion in oxidative conditions.

The combustion type is defined by the zone or area of the combustion process where the recovery or conversion of the cooking chemicals takes place. The oxidation of the organic material is completed in the later phases of the process so that the process is always completed in oxidative conditions using excess air.

The following subsections describe several of the commercially available processes for the combustion of spent sulphite liquor.

3.2.2 The MEI process

The MEI (Marathon Engineering Inc.) evaporation system incorporates direct contact evaporation with either waste incineration, fuel combustion, or a combination of both for evaporation heat requirements. The liquor can be concentrated for sale if the product market warrants.

Steam requirements for the MEI system are low, so additional boilers are not necessary. This is because the plant requires no steam for evaporation, only small amounts for vacuum eductors, fuel atomization, etc.

3.2.2.1 MEI liquor collection. Waste sulphite liquor is collected in non-agitated blow chests. The blown pulp is initially covered with stratified recycled liquor. Because of the density gradient achieved in the non-agitated blow chest, liquor concentration ranges from six percent (at the bottom) to one percent. Acid condensate and conventionally used whitewater follow in sequence. The process simulates counter-current washing and greatly improves spent liquor collection. On each blow, part of the recycle liquor is returned (by spraying) to the liquor collection tank. Under stable conditions, a one percent to six percent density gradient is established. Makeup for the recycle liquor is provided by the acid condensate used for blow pit pulp washing.

The acid condensate poses a problem; it has to have high acid content for effective scale removal in evaporator equipment, yet this also means high BOD_5 content in the post-wash condensate (non-recyclable) entering the mill's effluent. By controlling acid content in any one of a number of ways, the condensate, though no longer strong enough for rapid scale removal, is much cleaner, and can be used in acid makeup or pulp washing.

3.2.2.2 MEI evaporation. The MEI vacuum evaporator shown in Figure 15 is specially designed to minimize scaling caused by liquors containing inverse solubility salts such as calcium sulphate. In each unit there is a nucleation chamber situated immediately after heating and before evaporation. Thus, peak supersaturation conditions are achieved at the highest temperature. At this stage, the supersaturation is removed in the nucleator before evaporation. Because of inverse solubility, evaporative cooling brings the liquor back below its saturation level again before it

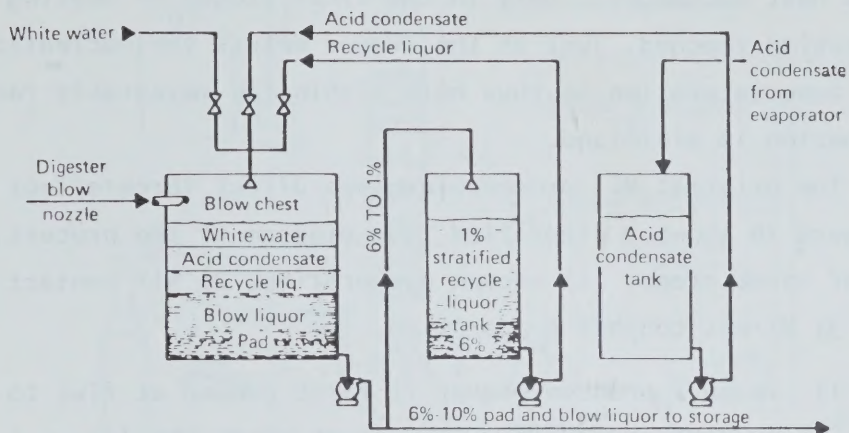


FIGURE 14. MEI BLOW PIT RECYCLE WASHING

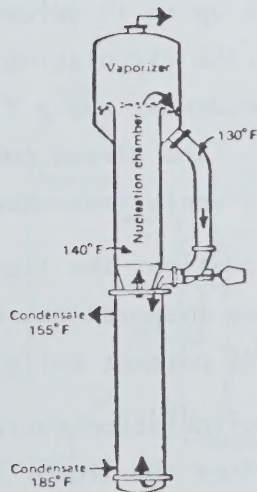


FIGURE 15. MEI REVERSE CYCLE SCALE CONTROL EVAPORATOR

enters the heat exchanger. Only in the final stages of heating is supersaturation reached, just as the liquor enters the nucleation chamber. Supersaturation is thus held within the metastable range, and scale formation is minimized.

The original MEI process involves direct three-effect evaporation. Figure 16 shows a simplified flow diagram of the process, which consists of three steps: 1) vacuum evaporation, 2) air contact evaporation, and 3) direct contact evaporation.

- 1) In this process liquor is first pumped at five to six percent solids to a flash tank where the liquor is exposed to a vacuum of 26 inches of Hg. Here volatile organics are stripped from the liquor.
- 2) Then the liquor proceeds to an air contact evaporator (cooling tower) where liquor is circulated through a packing through which air is simultaneously drawn. The air contact evaporator in this case operates just like a cooling tower common to most large scale refrigeration plants. While in the air evaporator, liquor is concentrated from about 8 percent up to 14 percent solids content. Heat is transferred to the circulating liquor in the vacuum evaporator, surface condenser and a flue gas temperature polishing heat exchanger. The direct contact evaporator, in essence, is similar to a venturi scrubber.
- 3) After air evaporation, the liquor is exposed to three stages of vacuum evaporation, which raises the concentration to about 55 percent solids.

In the MEI evaporators, the circulation cycle is reversed in order to minimize scale formation. After the liquor has been heated, it becomes saturated with calcium sulphate (a contaminate in the pulping liquors), a reverse solubility salt (comes out of solution at higher temperatures). With reverse cycle evaporators, immediately after heating, the liquor is introduced to a nucleation chamber where the liquor, saturated with calcium sulphate, forms small calcium sulphate crystals in the solution and thus, not on heat exchanger tubes.

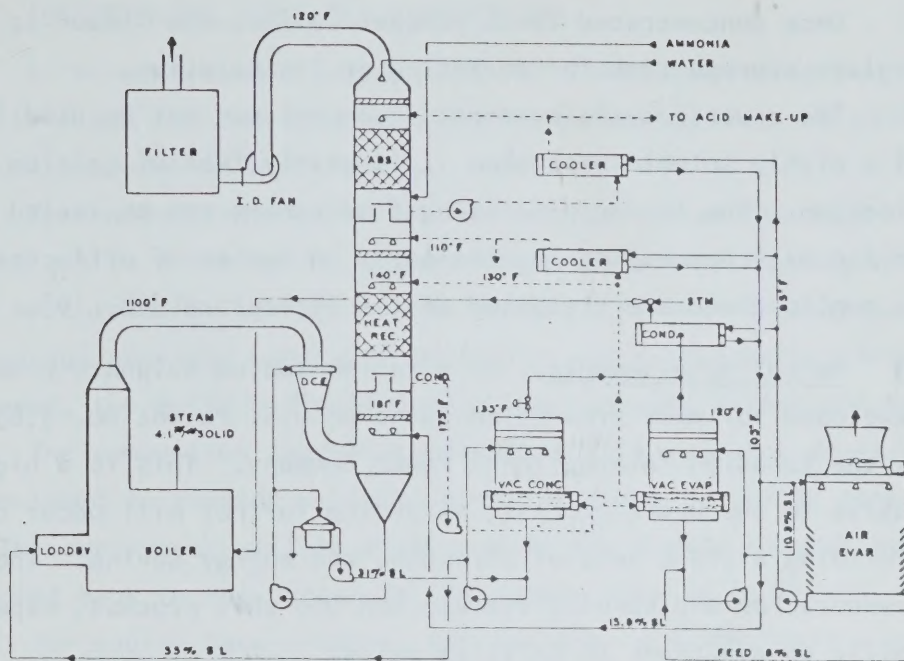


FIGURE 16A. SIMPLIFIED FLOW DIAGRAM OF MEI AMMONIA BASE
DIRECT THREE-EFFECT EVAPORATION AND BURNING SYSTEM

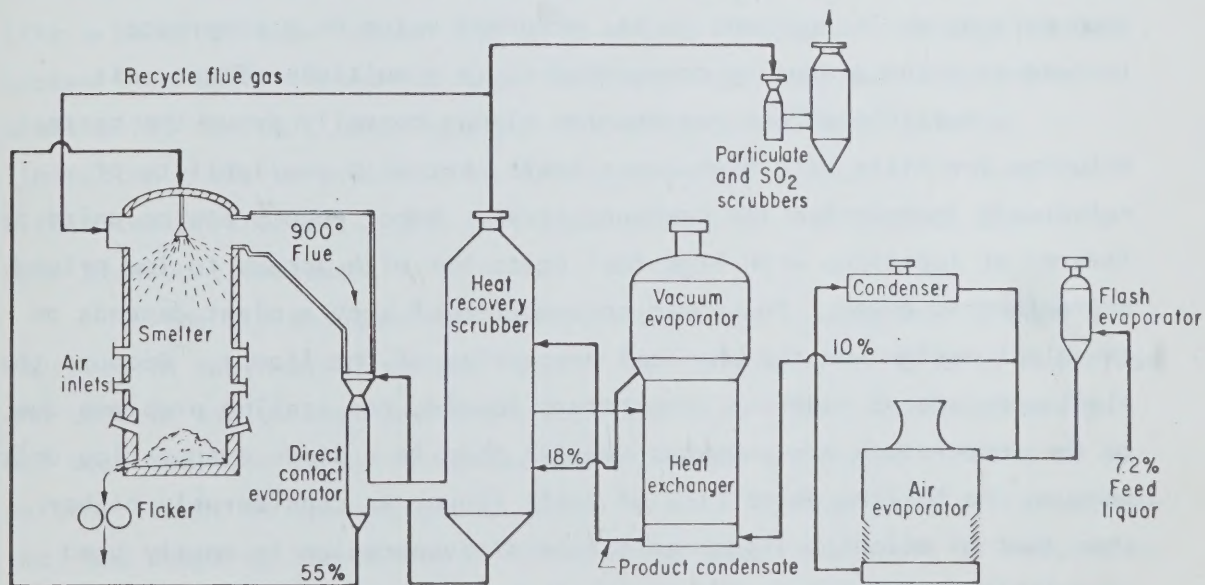


FIGURE 16B. SIMPLIFIED FLOW DIAGRAM OF SODIUM BISULPHITE
DIRECT THREE-EFFECT EVAPORATION AND BURNING SYSTEM

Once concentrated to 55 percent solids, the liquor is routed to a fiberglass storage tank for marketing or incineration.

The reverse cycle evaporator concept may not be used in the case of a highly soluble base when it is nearly free of calcium sulphate contamination. The liquor processing flow scheme can be varied according to specific mill processing requirements. A number of different specific process applications are discussed in the literature [18, 19].

3.2.2.3 Recent developments. An alkaline sodium sulphate process has been developed for mill production based on U.S. Patent No. 3,630,832, held by the Canadian International Paper Company. This is a high yield alternative to the kraft process, affording further mill odour control, lower refining without loss of strength, and energy saving. The MEI and other evaporation and burning systems can use this process, especially small mills or additions to existing mills.

3.2.3 Vapor recompression

In a vapor (or thermo) recompression plant, the pressure of the secondary steam is upgraded to its original value by a compressor, instead of being gradually downgraded as in a multiple effect unit.

Multiple effect evaporation plants normally prove the correct solution for mills with high power costs, but with availability of relatively inexpensive low pressure steam. Vapor recompression units are favored at locations with high fuel costs but with access to low priced hydroelectric power. The power consumption of such a plant depends on the plant design and the physical properties of the liquor. Because the plants operate at moderate temperature levels, the scaling problems due to the temperature are somewhat smaller than in a steam evaporation unit. Because the boiling point rise of kraft liquor is considerably higher than that of sulphite liquor this type of evaporation is mostly used in sulphite processes or for pre-evaporation in kraft operations.

Changing from multiple effect to vapor recompression plants does not basically influence stripping of volatiles and, thus, the secondary condensate BOD₅ load stays much the same. Also, evaporator body designs are similar.

3.2.4 Incineration after MEI evaporation

In direct contact evaporating for final concentration, any burner, fluidized bed reactor, furnace, or incinerator can be used without loss of evaporation efficiency.

The combustion chamber can be a large combustion separator to remove the ash to a collection system at the bottom of the separator. With the heat transfer surface of the recovery boiler eliminated, large clear openings provided, and with operation maintained below sintering temperatures, the buildup problems with the recovery boiler can be avoided. The combustion separator operating with very low density flue gas is designed to provide collection efficiencies in the 90 percent range. The carry-over will be picked up in the direct contact evaporator and recycled back to the burner in the liquor when it is burned.

For sodium base liquors, the furnace is essentially a refractory-lined smelter with the smelt collected and removed in the conventional manner. The smelt from a NSSC mill can also be cooled and solidified in a flaker, where the smelt is passed between chilled rolls and discharged like a sheet of paper about 1/32 inch thick. It is brittle and can be broken into small flakes for easy handling, storage, and shipment to kraft mills for salt cake supply. The MEI process does have the limitation that, at present, no working chemical recovery is possible within the mill itself. The salt cake market is limited and dependent upon transport distance.

The MEI designed smelter is a straight cylindrical design improved in many ways from the "Broby" type smelters previously installed and used in sodium base and kraft applications. The refractory design has been modified to avoid refractory expansion problems. The 50 to 60 percent liquor is sprayed on the walls high in the smelter, is concentrated, and drops to the bottom of the furnace where the smelt is reduced by the limited heated primary air. Secondary air is added higher in the smelter to control combustion in a manner similar to that of the recovery boiler and with the same excess air requirements (20 to 25 percent).

3.2.5 Sodium base recovery systems

3.2.5.1 SCA-Billerud. The SCA-Billerud process is based on pyrolysis of concentrated sodium base spent liquor at temperatures between 700°C

and 760°C (1292°F and 1400°F). It is unique in that no smelting operation is required. (Another pyrolysis process is the AST process; however, it never advanced from the pilot plant stage.) The almost instantaneous reaction is controlled to convert the sodium salts of the waste liquor into pure carbonate. At the same time, most of the sulphur is converted into hydrogen sulphide. This reaction takes place at temperatures below the melting point of sodium carbonate. The resulting reaction products, therefore, are a dry powder and a combustible gas. The powder is separated from the gas, and the sodium carbonate is leached out of the carbon residue by water.

The carbon is filtered off and returned to pyrolysis along with the waste liquor. By the combustion of the pyrolysis gases, the hydrogen sulphide is converted into sulphur dioxide in the flue gases. The last operation in the process is the reaction between the sulphur dioxide in the flue gases, and the sodium carbonate solution.

A flow diagram of the process used in the sulphite pulping is shown in Figure 17 and can be described, referring to Figure 17, as follows [20]:

1) The Reactor

The concentrated spent liquor is atomized by high pressure into the middle of the reactor, where it meets the hot flue gases from the oil burner at the top. The pyrolysis resulting from this encounter breaks down the organic compounds in the liquor almost immediately. The sulphur in the different sulphur compounds is converted into hydrogen sulphide and the sodium in the sodium compounds into sodium carbonate.

2) The Reactor Boiler

The reaction products from the reactor are gas and powder at high temperature. The gas contains hydrogen, carbon monoxide, carbon dioxide, hydrogen sulphide, methane, and nitrogen. The powder consists mainly of sodium carbonate, with some carbon and a small quantity of sodium sulphate. All the reaction products pass through the reactor boiler for heat recovery of the thermal heat as high pressure steam.

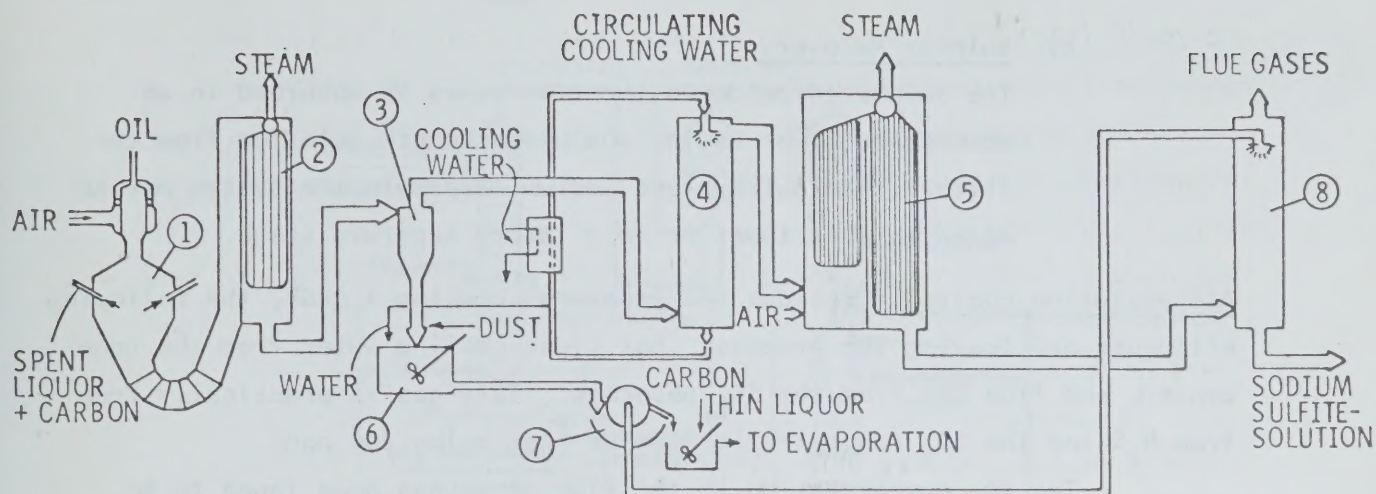


FIGURE 17. THE SCA-BILLERUD RECOVERY PROCESS IN SULPHITE PULPING

3) Powder Separation

The powder is separated from the gases in a two-stage cyclone separator.

4) Gas Cooling

In order to enrich the gases before burning, the wet pure gas is cooled in a scrubber tower for water vapor condensation.

5) Gas Fired Boiler

The cooled pyrolysis gas is burned, and the heat from the combustion is recovered as high pressure steam in this common type boiler.

6) Leaching

The powder from the cyclones is mixed with water and the sodium carbonate and sodium sulphate are leached out of the carbon residue.

7) Carbon Filtration

The sodium carbonate solution mixed with carbon is filtered. This yields a pure sodium carbonate solution, which goes to the absorption tower for sulphur recovery. The remaining carbon is mixed with spent liquor and returned to the reactor.

8) Sulphur Recovery

The sulphur dioxide in the flue gases is absorbed in an absorption tower by the sodium carbonate solution from the filter. The sulphur and sodium makeup to the system may be added in this tower or in a later, separate stage.

Air pollution control - Besides the recovered cooking liquor, the following effluents are leaving the process: hot clean cooling water from the condenser, and flue gas from the SO_2 absorber. This gas is practically free from H_2S and the SO_2 concentration can be kept below 200 ppm.

The NO_x concentration in the flue gases has been found to be in the range of 10×20 ppm.

Chemical and energy recovery - A plus with this system is a possible positive return on investment from the chemical and the heat recovery savings, rather than a negative or minus environmental cost (if effluents and emissions were tackled independently).

In other words, since many sulphite mills have to either shut down or stop emissions, the cost for normal solutions would be charged against environmental clean up. This system though may actually give a payback which, of course, varies with the cook and size of the mill operation.

3.2.5.2 Stora. Figure 18 shows the simplified flow diagram for the sodium base Stora sulphite chemical recovery process. The key principles of the process are as follows:

The clarified green liquor from the recovery boiler, containing sodium sulphide and sodium carbonate, is carbonated with recirculated carbon dioxide. Sodium sulphide is thereby converted to hydrogen sulphide that is stripped from the liquor by using carbon dioxide as carrier, and is brought to a "Claus Reactor" where the hydrogen sulphide reacts with sulphur dioxide to form elemental sulphur, the carbon dioxide is recirculated within the process. Thus, the green liquor is converted to sodium bicarbonate. This is then reacted with sodium bisulphite to produce sodium sulphite. Part of the sulphite solution absorbs sulphur dioxide in an absorption tower and is returned as bisulphite.

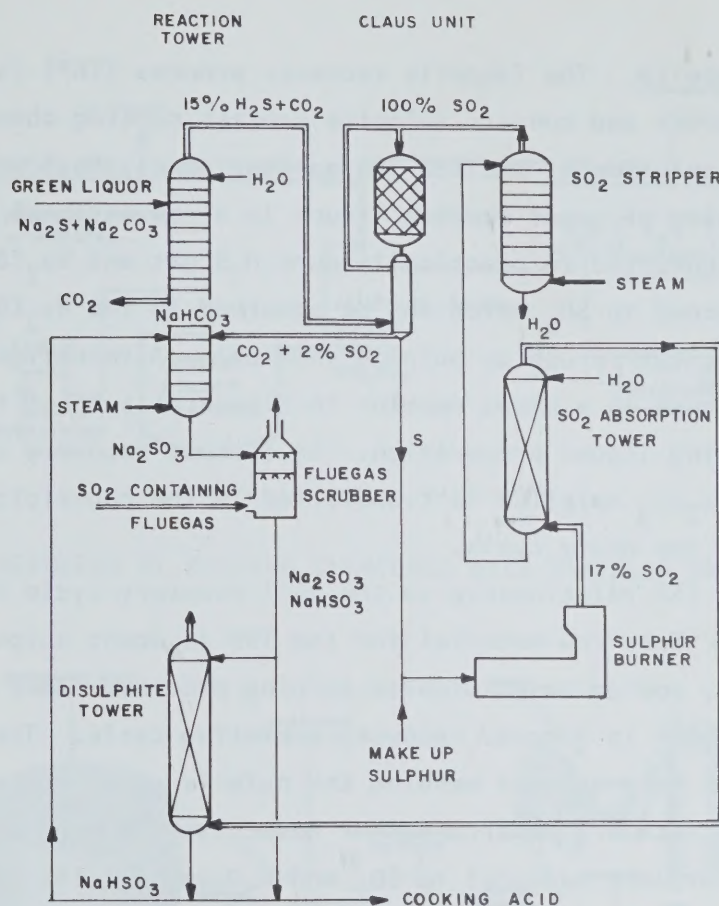


FIGURE 18. STORA CHEMICAL CONVERSION PROCESS

Depending on process requirements, the cooking liquor is made from suitable proportions of the sulphite and bisulphite solutions. When higher pH is needed, as for semi-chemical pulping, some bicarbonate is bypassed to the sulphite solution.

Energy-requirements - Steam and power demand is low. For a conversion plant serving a 400 tons per day mill making a two-stage cooked paper grade pulp, the steam and power demand for the conversion plant would be approximately as follows:

Power: About 300 kW; of this 50 kW is for the Claus heaters and 75 kW for the sulphur burning system.

Steam: Four to five tons per hour; of this three to four tons per hour is for the SO₂ stripping. If the sulphur burner has a recovery boiler this will produce about four tons per hour, leaving a net requirement of less than one ton per hour.

3.2.5.3 Tampella. The Tampella recovery process (TRP) is a chemical system to recover and convert sulphite process cooking chemicals. The green liquor solution of smelted inorganics, namely Na_2S and Na_2CO_3 from the burning of spent cooking liquor in a conventional kraft recovery furnace, is converted to practically pure H_2S gas and Na_2CO_3 solution. The H_2S is burned to SO_2 which can be absorbed in the Na_2CO_3 solution to form Na_2SO_3 for re-use as pulping chemical. Alternatively, the H_2S can be burned in a Claus reactor to elemental sulphur for polysulphide cooking liquor preparation. In a cross recovery application, part of the Na_2CO_3 solution is transferred to the causticizing department in the draft cycle.

The TRP relationship to the mill recovery cycle is illustrated in Figure 19. Base raw material for the TRP is spent sulphite liquor (SSL) from any sodium base sulphite pulping process. This is blended with kraft liquor in a cross recovery operating cycle. The weak spent cooking liquor recovered in washing the pulp is concentrated and transferred to the recovery department for reductive burning, with subsequent oxidation of organic material to CO_2 and H_2O vapor. The sulphur compounds are mainly reduced to Na_2S , and the balance of the sodium forms Na_2CO_3 . The furnace may be of the conventional kraft recovery boiler design or a smelter for burning with minimum heat recovery. The smelt of Na_2S and Na_2CO_3 taps from the furnace into the green liquor dissolving tank for dissolution.

Economics of TRP establish that the green liquor concentration of total alkali should be 180 g of Na_2O per litre. The green liquor to be processed in the TRP must be clarified. Improved clarification of green liquor can be accomplished by using lime and flocculating agents.

3.2.5.4 Fluidized bed incineration. Fluidized bed recovery systems for the sulphite pulping processes have many features in common. The systems discussed here are basically designed for magnesium and ammonium base mills but are also applicable to sodium acid sulphite and calcium base. Figure 20 shows an example of the fluidized bed system for acid sulphite and bisulphite recovery.

Weak spent liquor received from the pulp mill washing system passes through storage to the multiple effect evaporators for evaporation

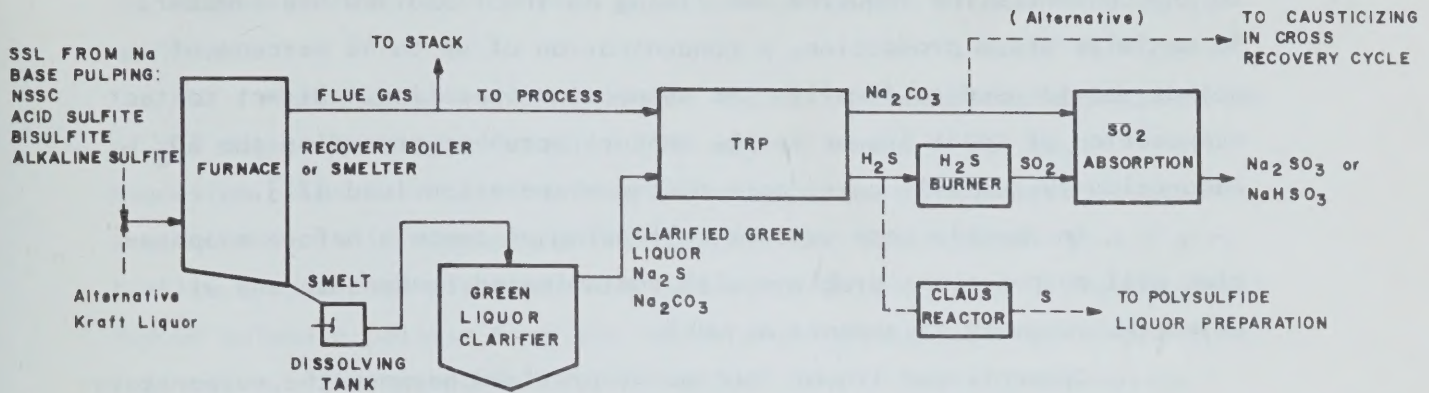


FIGURE 19. RECOVERY OF PULPING CHEMICALS WITH TAMPELLA RECOVERY PROCESS

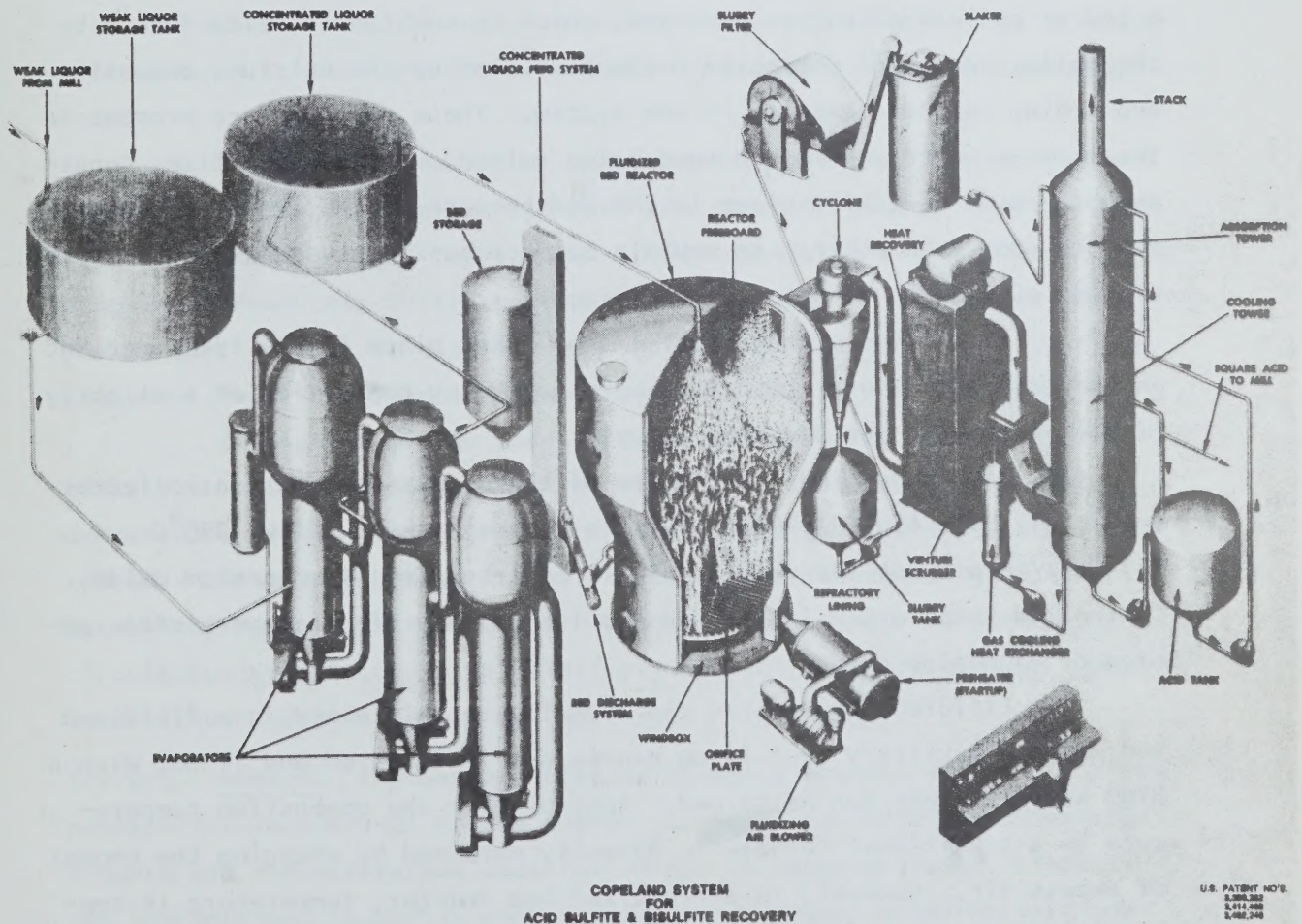


FIGURE 20. COPELAND FLUIDIZED BED SYSTEM FOR ACID SULPHITE AND BISULPHITE RECOVERY

to the concentration required for firing in the fluidized bed reactor. To maximize steam production, a concentration of up to 46 percent of solids may be used to simplify the evaporator operation. Direct contact evaporation of spent liquor in the venturi scrubber preceding the SO_2 absorption system will carry part of the evaporation load [21].

In ammonia base systems, stripping of ammonia before evaporation will help prevent problems with contaminated condensate and will allow the recovery of ammonia as well.

Concentrated liquor storage is provided between the evaporators and the combustion system. This provides buffer capacity and allows for continuous operation of the combustion system during periods of evaporator boilout. Liquor is injected into the freeboard of the fluidized bed reactor through a single feed gun. The fluidized bed reactor operates on a bed of pelletized magnesium oxide, which is constantly being formed by the pelletization of magnesium oxide particles on the calcium, potassium and sodium sulphate present in the system. These elements are present in the process water and in the wood being pulped. At the same time, constant attrition is taking place in the bed because of the impact of the pellets upon each other. In ammonia base combustion systems the bed will consist of the ash residual of the liquor.

Combustion of the spent liquor takes place in the freeboard and in the bed, with the freeboard temperature being controlled at a slightly higher level than the bed temperature.

The combustion temperature of the process can be controlled at any level above 650°C (1202°F), but is purposely kept below 1090°C (1944°F) to prevent the dead burning or overburning of magnesium oxide. Controlled temperature in the bed results in very high recovery efficiencies of magnesium oxide.

Efficient combustion can normally be maintained in a fluidized bed without auxiliary fuel. Low excess air is required and liquor with a high water content can be burned. Control over the combustion temperature in a traditional furnace is normally obtained by changing the amount of excess air. However, in a fluidized bed reactor, temperature is controlled by maintaining a high water content in the spent liquor in the fluidized bed reactor. By minimizing the generation of sulphur trioxide,

the recovery efficiency of both magnesium oxide and sulphur dioxide is maximized.

The combustion temperatures result in the thermal decomposition of the magnesium sulphur complex into magnesium oxide and SO_2 . Magnesium oxide is recovered from the exhaust gas stream by cyclone separators. It is discharged from the cyclone separators into a slurry tank to be slurried with water and washed on a belt filter. Washing will remove sulphate ballast from the system and recycle pure MgO to the pulping system. The washed MgO is slaked and sent to the SO_2 absorption system.

The absorption section is a four-stage moving bed absorber, in which magnesium hydroxide slurry, diluted to the proper concentration with cooling section condensate, is in counter-current flow with the exhaust gases. An acid with a pH of 4.5 to 5 will be generated here and will be used in the fortification tower to generate the final cooking acid.

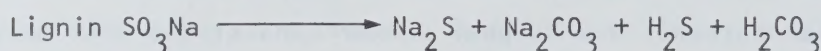
The required makeup of magnesium hydroxide will come from a magnesium hydroxide supply system and the required sulphur dioxide makeup from a sulphur burner. The sulphur burner exhaust gas will be cooled and directed through the fortification tower. The tail gas from the fortification tower can be sent through the absorption section of the main gas treatment tower to minimize sulphur dioxide losses.

Combustion of the spent liquor at low solids concentrations greatly simplifies and improves the evaporator operation. Three body evaporators with an equal number of effects are usually employed in the system. Continuous boilout is not required and the normal interval between boilouts is three to four weeks. By operating the forced circulation evaporators at a relatively low solids concentration, the spent liquor does not scale as readily.

3.2.5.5 Sonoco. The Sonoco Sulphite Recovery Process (SSRP) is a new process for recovery of sodium and/or sulphur from sulphite pulping waste liquors and reconstituting them for re-use as cooking liquor. Burning of the concentrated liquor is done in the solid state. Aluminum complexes are involved in the recovery chemistry and process.

The process was originally developed specifically for the Sonoco Products, Hartsville, South Carolina Mill which produces mainly corrugating medium. The process was subsequently licenced to other manufacturers and distributors. In addition to the Hartsville installation, which is now operational, the process is presently under construction at the mills of Groveton Paper Company, Groveton, New Hampshire and SCL Meat and Paper Pty. of Melbourne, Australia.

In the process, waste liquor is burned in a reducing atmosphere in the presence of hydrated alumina to yield sodium aluminate ash, carbon dioxide, and hydrogen sulphide. At the same time, the hydrogen sulphide formed is oxidized to sulphur dioxide in the same reactor. The sulphur dioxide is then reacted in a solution of the sodium aluminate ash to yield hydrated alumina for recycling to the process feed, and sodium sulphite for makeup chemical in NSSC pulping. The basic chemistry of the process involves the following possible reactions [22]:



Process description - In the process, which is outlined schematically in Figure 21, the spent NSSC liquor is concentrated in multiple effect evaporators to about 40 to 50 percent solids. The concentrated spent liquor is mixed with recycled finely divided reactive alumina hydrate in an amount necessary to convert the soda content of the liquor to sodium aluminate. This mixture is fed to a rotary pelletizer where sufficient finely divided pulverized recycled sodium aluminate ash is introduced to form more or less round, solid pellets as a result of the rolling action of the mass. The pellets may range in size from about 1/4 to 1 inch or

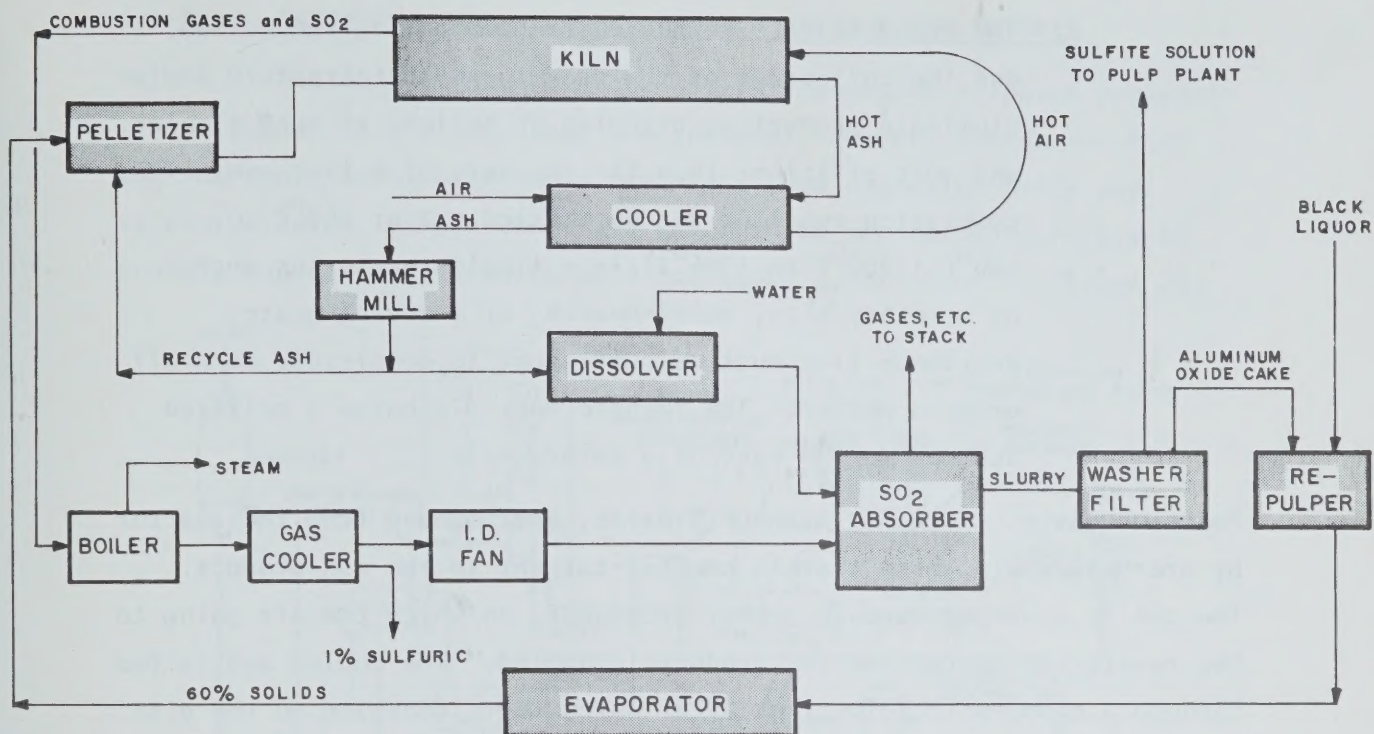


FIGURE 21. SONOCO SULPHITE RECOVERY PROCESS

more in diameter. The quantity of dry sodium aluminate required to form discrete pellets depends on the water content of the spent liquor feed. At 50 to 60 percent solids, approximately equal parts by weight are used. There is considerable heat evolved in the pelletizing step through an apparent dehydration-hydration reaction, and the pellets are very cohesive and hard in spite of their water content. Proper pelle formation also is dependent on feed rate and rotation spent of the pelletizer.

The pellets serve two very important functions:

1. In the pellet form, the liquor solids are encapsulated with the alumina reactant within the pellet. Accordingly, during combustion, a reducing atmosphere is inherently maintained for the necessary basic reactions, yielding sodium aluminate, CO_2 , and H_2S , regardless of the combustion conditions outside of the pellets. In the oxidizing atmosphere outside of the pellets, the H_2S is oxidized to SO_2 .

2. The round pellets retain their form during combustion, and the collection of the gray to white refractory sodium aluminate product as granules or pellets is much simpler and more efficient than the recovery of a fine ash. This combustion reaction can be carried out at 650°C to 980°C (1202°F to 1796°F) in a simple reactor or such as a rotary kiln, multi-hearth, or a moving grate. Residence time must be sufficient to completely burn all organic matter. The furnace must discharge a pelleted ash, as in the case of a rotary kiln.

Reaction gases, including sulphur dioxide, are removed from the reactor by pre-heated air which travels counter-current to the ash product.

The ash is conveyed through a heat exchanger, in which the air going to the reactor is heated and the product is cooled. The cooled ash is fed through a hammermill pulverizer after which it is conveyed to the dissolver tank, where it is mixed with water to form a solution of sodium aluminate. The solution is prepared in a concentration corresponding to the desired concentration in the final product.

Gases from the reactor are conveyed by fans through a cyclone and heat recovery boiler which produces 150 psi steam, then through a gas cooler, where it is cooled to about 70°C (158°F). Ash fines from the cyclone are conducted to the dissolver tank.

The ash solution and the off-gases containing sulphur dioxide are contacted counter-currently in an absorber. The pH value is lowered here so that aluminum hydroxide is precipitated from the aluminate solution and soluble sodium sulphite is produced. If insufficient sulphur dioxide is present, some carbonate will also be formed.

The slurry is then passed over a filter washer system where the hydrated alumina cake is removed. The filtrate, primarily a sodium sulphite solution, is pumped to the pulp plant for use in making cooking liquor.

The hydrated alumina cake is mixed with incoming raw liquor which is enroute to the pelletizer to be incorporated as part of the feed to the reactor.

3.2.6 Magnesium base recovery using a furnace and boiler

Residual liquor from the several magnesium sulphite processes can be burned in a simple heat and chemical recovery system in which the inorganic salts break down precisely into magnesium oxide and sulphur dioxide. These chemicals can then be recombined directly to produce magnesium bisulphite acid for cooking, as shown in Figure 22.

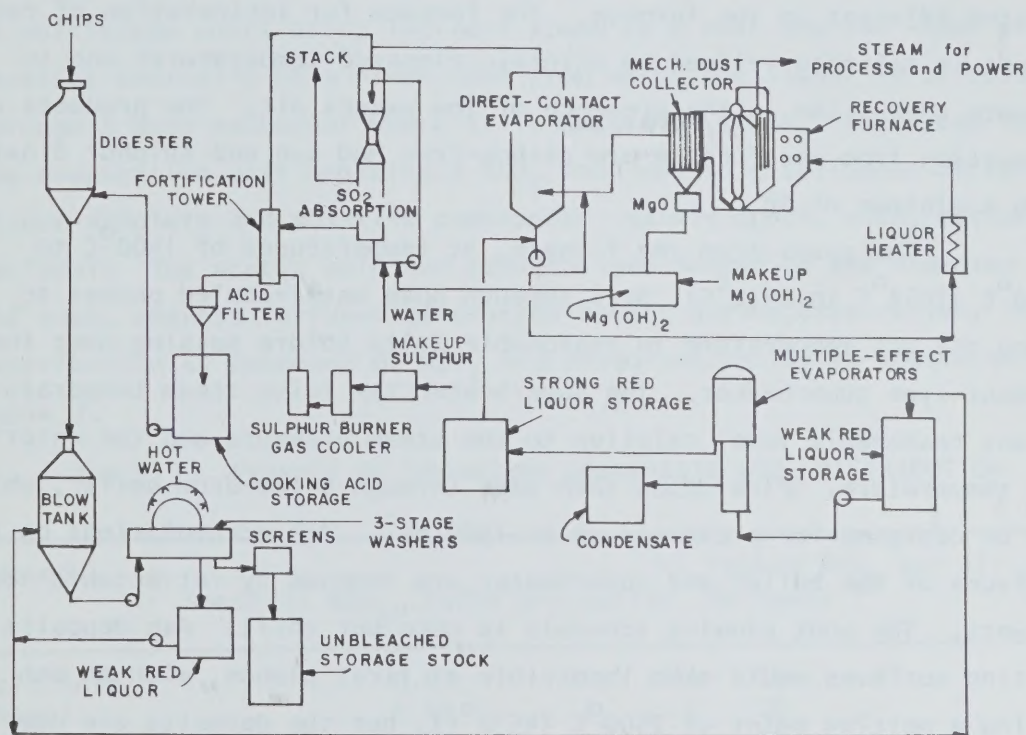


FIGURE 22. FLOW DIAGRAM OF MAGNESIUM BASE AND RECOVERY

3.2.6.1 Liquor incineration. The residual liquor (red liquor), having been evaporated to the required concentration, is fired into a furnace through circular burners similar to oil burners. The liquor burns in suspension in an oxidizing atmosphere with excess air as low as is compatible with complete combustion. Excess air is minimized to avoid the formation of magnesium sulphate in the ash. The circular liquor burners are equipped with sprayer plates where, in combination with steam, the liquor is atomized to facilitate rapid and complete combustion. The burners are also located in opposed positions in the side walls of the furnace, thereby providing turbulence to ensure complete combustion in the atmosphere of low excess air.

The air to support liquor combustion is introduced through the circular burner coincident with the liquor flow. To ensure rapid and complete combustion of the liquor, it is desirable to have the air at temperatures in the range of 600°F to 700°F (315°C to 371°C). These temperatures can be obtained with a tubular air heater located in the path of the combustion gases or with a direct fired air heater located adjacent to the furnace. The furnace for incineration of red liquor is refractory-lined to maintain elevated temperatures and to promote combustion in the presence of low excess air. The products of combustion from the furnace are carbon-free MgO ash and sulphur dioxide with a minimum of SO₃.

The gases from the furnace, at temperatures of 1400°C to 1600°C (2552°F to 2912°F), pass through open water-cooled passes to bring the gas temperature to reasonable limits before passing over the pendant-type superheater. The superheater can raise steam temperatures to any reasonable level relative to the steam pressure and the entering gas temperature. Flue gases then pass through a two-drum boiler, which can be designed for pressures up to 1500 psi. Ash accumulations on the surfaces of the boiler and superheater are removed by retractable soot blowers. The soot blowing schedule is once per shift. Ash deposits on heating surfaces would seem impossible at first glance, with an ash having a melting point of 2500°C (4532°F), but the deposits are due to contamination by sodium and potassium derived from the wood ash, water, and chemicals.

The magnesium base recovery system described herein can be used for any magnesia base pulping technique. Coincidentally with chemical recovery, heat value in the residual liquor is recovered as steam for power and process. It is difficult to draw direct comparisons of gross steam per ton of pulp because of the influence of process yield, wood species, and grade of pulp. However, thermal efficiency of MgO recovery is five to eight percent higher than soda base recovery systems because of lower ash and sulphate content in the liquor, which reduces the heat reaction and the heat reduction required [23].

The sulphite industry in Canada generally is seeking to produce as high yield a sulphite pulp as possible for chemical furnish to newsprint

machines. A magnesium base pulp cannot be made at as high yield as can sodium base pulp. This is important in a mill's selection of the pulping base and chemical recovery cycle.

3.3 Evaporator Condensates

3.3.1 Chemical characteristics

The concentration of spent sulphite liquor is usually performed in multistage units using indirect steam as a heat source. Each stage consists basically of a flash tank from which the liquor is circulated through a heat exchanger where it is heated by steam. The steam leaving the evaporation unit contains a BOD_5 load due to entrainment of spent liquor droplets and volatile compounds: mainly acetic acid, methanol, and furfural. The acetic acid and methanol are formed in the digester during the cook, whereas furfural generation occurs during evaporation. The distribution by compound of BOD_5 and dissolved solids is presented in Table 7.

TABLE 7. EXAMPLE OF SECONDARY CONDENSATE BOD_5 DISTRIBUTION

Compound	Specific BOD_5 , lb/lb dry solids	Pounds BOD_5 per air dried ton	
		Hardwood	Softwood
Acetic Acid	0.67	60	29
Methanol	1.00	26	24
Furfural	0.85	13	6
Miscellaneous		4	2
Total		103	61

Due to the high volatility of methanol and furfural, about 95 to 100 percent of the amount entering the evaporation plant is found in the secondary condensate. About 50 to 70 percent of the acetic acid entering with the spent liquor is also found with the secondary condensate. The distillation factor of acetic acid is dependent on the pH of the spent liquor, as shown in Figure 23 [13].

The actual carry over of BOD_5 in the form of spent liquor droplets depends on the physical layout and operating parameters of the

unit in question. The installation of mesh pads or any other type of droplet separating device in the steam/liquor separator reduces the liquor carry over, but also introduces an additional pressure drop in the system.

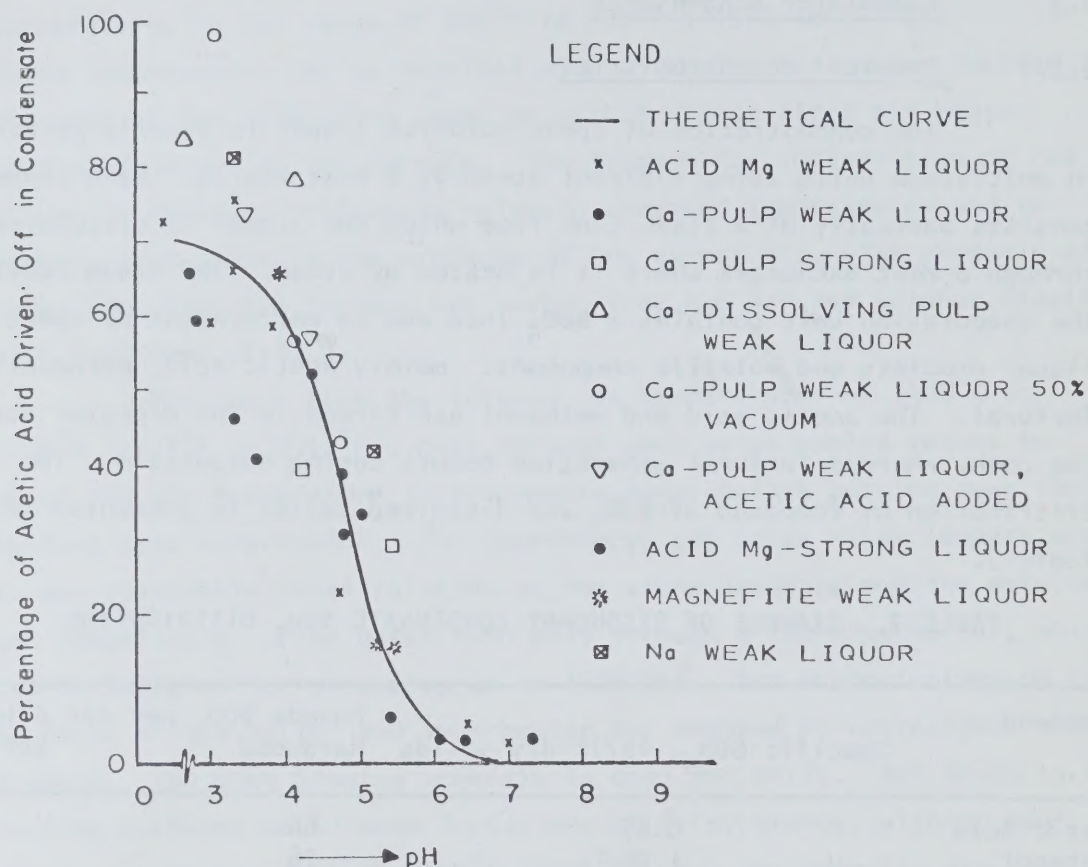


FIGURE 23. THE INFLUENCE OF SPENT SULPHITE LIQUOR pH ON THE PERCENTAGE OF ACETIC ACID DRIVEN OFF IN CONDENSATE

3.3.2 Recycling of condensates

About 30 percent by volume of the condensate can be returned to the acid preparation plant [24]. However, if the fractions richest in methanol and furfural are chosen, the greater part of these pollutants can be trapped. Further, condensate re-use in the flue gas scrubber or in recovering SO_2 from flue gas will result in only 30 percent of the furfural and methanol remaining in the condensates. Alternatively, the furfural and methanol can be stripped from the condensates with steam in a special column [25].

3.4 Energy Consideration

A concern exists in Canada that chemical pulp mills be made as energy self-sufficient as practical. In some cases, the cost of generating or recovering energy may be as expensive as off-site purchase of energy. However, it may be necessary to follow this concept because of the expected tight energy supply [26].

3.4.1 Process energy demands

Table 8 shows the ratio of power and heat demand of some typical wood based processes. The table also presents the process heat generation. This relative comparison is based on a process heat demand in the form of steam expressed as 100 percent. The power demand includes all the energy supplied in the form of electricity. Process heat generation expressed as the heat of steam includes the heat of waste liquors, bark, sawdust, grinding powder and similar process wastes.

TABLE 8. RELATIVE ENERGY DEMAND AND GENERATION OF WOOD BASED PROCESSES*

Product or mill	% Power Demand	% Process Heat Generation
Softwood kraft pulp - unbleached, dried	17	120
Softwood kraft pulp - bleached, dried	20	110
Birch kraft pulp - bleached, dried	21	105
Eucalyptus kraft pulp - bleached, dried	19	80
Sulphite pulp, Ca-base - unbleached, dried	17	95
Sulphite pulp, Mg and Na-base - unbleached, dried	18	100
Groundwood pulp - flash dried	230	50
Thermo-mechanical pulp - flash dried	290	50
Tissue paper (woodfree) mill	60	0
Printing paper (woodfree) mill	45	0
Newsprint (+ groundwood) mill	85	15
Kraftpaper (+ sulphate pulp) mill	30	105
Printing paper (+ sulphate pulp) mill	30	80
Sawmill	13	115
Fiberboard mill	20	0
Particle board mill	30	30

* Based on process heat demand = 100%

The figures in Table 8 indicate that the power demand in chemical pulping is low compared to the heat demand, and also that process heat generation is almost equal to process heat demand. Considering that the possible byproduct power generation is about 15 to 20 percent calculated on the process heat demand, or about the same quantity as the power demand, it can be concluded that the chemical pulping processes can be very nearly self-sufficient in heat and power [27].

3.4.2 Pulp mill energy balance

The heat balances for a softwood kraft mill are presented in Figure 24. The heat balance for an unbleached magnafite pulp mill has been added for the sake of comparison. The left bar represents the heat demand of process and power generation, the right one the heat generation with the process fuels (black or waste liquor and bark) and purchased fuel if needed. Heat generation with sulphite waste liquor is slightly lower compared to sulphate black liquor. Thus, the amount of additional heat in the form of purchased fuel is about 15 percent of the total heat demand.

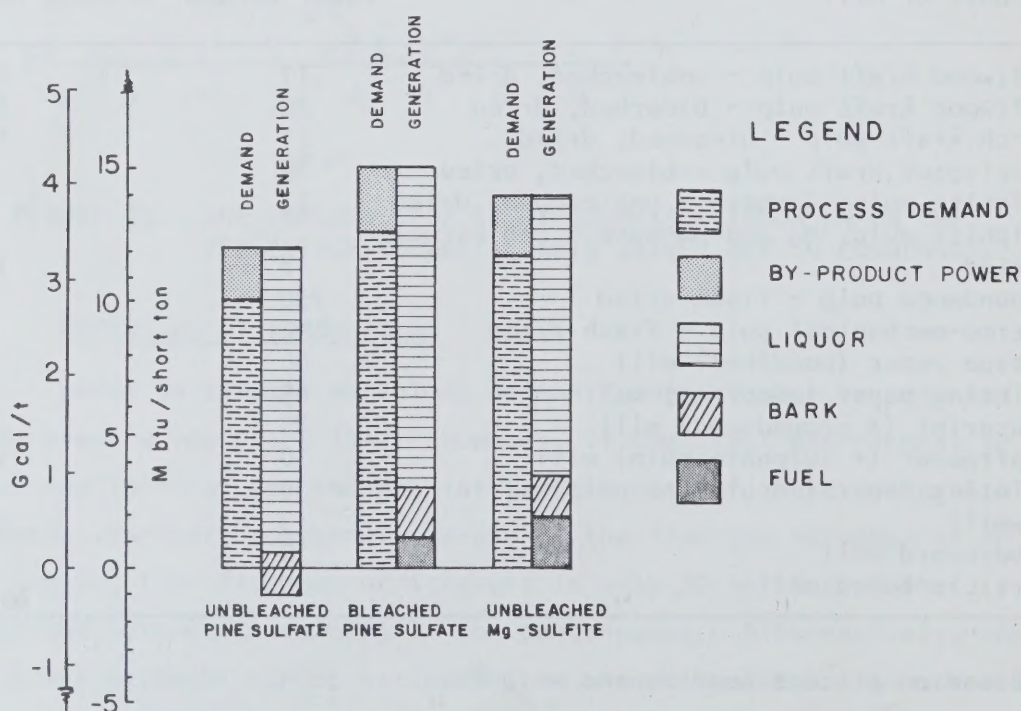


FIGURE 24. PULP MILL HEAT BALANCES FOR VARIOUS COOKING PROCESSES

Power balances for the same pulping processes are presented in Table 9. All of the processes have a fairly high degree of self-sufficiency and only 10 to 20 percent of the power has to be purchased. No essential possibilities to increase byproduct power production can be expected. On the high pressure side, recovery boiler corrosion limits the increase in the back pressure turbine enthalpy difference, and significantly lower process steam pressure can hardly be used.

TABLE 9. PULP MILL POWER BALANCE

	Unbleached Sulphate (kWh/t)	Bleached Sulphate (kWh/t)	Unbleached Magnesite (kWh/t)
By-product Power	550	700	600
Purchased	60	145	100
Generation = Demand	610	845	700

3.4.3 Economics of spent sulphite liquor recovery

A major factor in the economics of sulphite pulping is the estimated difference between cost incurred with recovery and no recovery. This economic analysis has recently been conducted considering both the ammonia sulphite and magnesium sulphite pulping processes, each of which has different heating values [28].

This analysis has shown that smaller mills and mills with higher cooking yields (65 percent) will have poorer economic results, with half the net heat available as that for paper grade pulp. Dissolving grade pulp gives a better result than paper grade pulp. The higher the fuel price is, the more attractive the economics of liquor recovery are. An oil price of about \$15 per barrel is required to give a positive return on investment (ROI) for a 500 air dried ton per day ammonia base mill. Under current conditions, normally the recovery system costs money to operate and has a negative ROI. Table 10 shows the effect of mill capacity on ROI for the ammonia base mills in the 100 to 600 air dried ton per day (ADT/day) range at an oil price of \$10 per barrel.

TABLE 10. THE EFFECT OF MILL CAPACITY ON ROI, AMMONIA PROCESS, OIL PRICE \$10/BARREL

	Mill Capacity (adt/day)					
	100	200	300	400	500	600
90% Collecting Efficiency Investment Million \$	10.6	16.2	20.5	24.6	27.9	31.1
Fixed Costs (17% of Investment) Million \$/yr	- 1.8	- 2.8	- 3.5	- 4.2	- 4.7	- 5.3
Credits from Oil Savings Million \$/yr	+ 0.7	+ 1.3	+ 2.0	+ 2.7	+ 3.3	+ 4.1
Operation and Maintenance Costs Million \$/yr, (Minus returned sulphur)	- 0.2	- 0.2	- 0.2	- 0.2	- 0.3	- 0.3
Net Cost Million \$/yr	- 1.3	- 1.7	- 1.7	- 1.7	- 1.7	- 1.5
ROI %	-12.0	-10.0	- 8.0	- 7.0	- 6.0	- 5.0
Increasing Cost \$/ADT	37.0	23.0	16.0	12.0	10.0	7.0
95% Collecting Efficiency Investment Million \$	11.3	17.2	21.8	25.9	29.5	33.0
Fixed Costs (17% of Investment) Million \$/yr	- 1.9	- 2.9	- 3.7	- 4.4	- 5.0	- 5.6
Credit from Oil Savings Million \$/yr	+ 0.7	+ 1.4	+ 2.1	+ 2.8	+ 3.5	+ 4.3
Operation and Maintenance Costs Million \$/yr (Minus returned sulphur)	- 0.2	- 0.2	- 0.2	- 0.2	- 0.3	- 0.3
Net Increasing Costs Million \$/yr	- 1.4	- 1.7	- 1.8	- 1.8	- 1.8	- 1.6
ROI %	-12.0	-10.0	- 8.0	- 7.0	- 6.0	- 5.0
Increasing Costs \$/ADT	40.0	24.0	17.0	13.0	11.0	8.0

+ Denotes Credit

- Denotes Cost

3.4.4 Other energy savings

During the past, power consumption by the pulp and paper industry has gradually increased by 2 to 3 percent per year. Part of this increase is caused by quite natural reasons: the shortage of manpower, better control of the process, and new requirements in environmental protection. However, part of the increasing power consumption is unnecessarily caused by the over-dimensioning of equipment, mainly of pumps. Pulp and paper production involves extensive pumping of low consistency stock flows and water from tank to tank. The pumps are mostly driven directly by a motor, tank levels being automatically controlled by control valves. Over-dimensioning cannot be verified as easily as under-dimensioning because the control valves reduce pressure and flow so that no disturbances occur in the process.

Improved equipment application can contribute to less power use. Most high yield unbleached sulphite pulp production, for example, has been based on two-stage disc refining of the high yield chips at a low (and inefficient) consistency of 6 to 8 percent. Recently [29], refining has been carried out in a single-stage of double disc refining at high consistency. This is accomplished in high power, high capacity refiners which permit separation by fiber to fiber action. These installations require one-third the power normally used in conventional atmospheric disc refiners for a pulp in the 75 to 80 percent yield range.

4. COST CONSIDERATIONS

4.1 Approach to Basis of Cost Consideration

Effluent control regulations require pulp mills to undertake measures either to reduce the amount of effluent polluting loads internally or to treat these effluents externally. Obviously this decision is largely economic, while realizing a technically acceptable solution.

While the cost for internal recovery basically involve the capital and operating cost of the additional equipment, the effect of such measures on the total system must also be considered. For instance, higher spent liquor collection efficiency achieved by changing the equipment and operation of the washing plant will affect the whole recovery cycle. In new mills, the additional capacity is available at moderate marginal cost of the equipment. In existing mills, the problem is more complicated because of the limited capacity of the equipment and the high initial price of additional capacity.

Mills with a low degree of liquor recovery will generally find that investment in internal recovery is more economically sound than investment in nonprofitable external treatment. An external secondary treatment plant will consume more net energy per unit of BOD_5 removed than internal recovery will. Toxic compounds in the SSL are destroyed by incineration via internal recovery. Secondary treatment has not been proved to produce a nontoxic effluent, which is a Canadian requirement.

Seven different spent liquor alternatives were selected for study as case examples. The mill production, pulping base, and yield are considered to be representative of existing or projected mill conditions for sulphite mills in Canada. A brief summary of these case examples is given in Table 11.

This section of the report considers capital and operating costs for each of the different case examples. In addition, information is provided on:

- (a) equipment description of the evaporator, burner, and chemical recovery;
- (b) percentage recovery of chemicals from the feed liquor;
- (c) power (kW/adt) and steam (lb/adt) requirements;

TABLE 11. SUMMARY OF SPENT LIQUOR RECOVERY CASE EXAMPLES

Case Example	Alternative	Mill Size tpd	Pulping Yield (%)	Recovered SSL Solids lb per adt	Feed Liquor to Evaporators (%)	Liquor Concentration after Evaporation (i.e. to burning) (%)
1	Sodium base bisulphite	120	70	1080	8.0*	52
2	Sodium base bisulphite	250	70	1080	8.5	52
3	NSSC	250	80	700	9.0	52
4	NSSC, NH_3	250	80	700	8.0	52
5	Magnesium	250	50	2040	9.0	52
6	Magnesium	250	70	1080	10.0	52
7	NH_3 -dissolving	500	44	2540	8.0	52

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* Feed liquor to evaporators = 8.5% for MEI System, Table 12.

- (d) manpower (operation and maintenance) per shift; and,
- (e) water (gallons/adt).

All costs have been adjusted to represent conditions for a mill located in Quebec with price levels for June, 1976. The cost information is mainly from equipment suppliers, previous experience, and information on file. All costs are quoted in Canadian dollars [30, 31]. When cost data adjustments for time were necessary, it was accomplished by using Engineering News Record cost indexes for Montreal, which is a standard engineering approach for adjusting cost to a common base time period and geographical location. An operating year was assumed to be 350 days. Because on-site land value would vary widely in practical cases, it was not included in the cost information. These examples also assumed the addition of recovery equipment to an existing facility and, for practical reasons, do not include any additional cost for replacement which may be incurred in removing old or otherwise inadequate existing equipment [32].

Obviously, such cost conditions can only reflect average conditions, and they can have no greater accuracy than normally expected for preliminary estimating purposes. In using and comparing these figures it must be observed that they are based on a set of hypothetical conditions and, as such, never directly apply to actual projects.

4.2 Liquor Recovery Cost

The cost of spent sulphite liquor (SSL) recovery depends on the solids produced rather than the air dried tons of pulp produced per day. Mill pulp production capacity is not a good function for cost comparison unless the yield is specified. The cost information presented in this section specifically refers to a spent sulphite recovery system in which the liquor is fed to the evaporators at a specified feed concentration. The systems considered herein do not include equipment for liquor washing or other ancillary equipment prior to evaporation and recovery.

The cost information referred to herein represents the items of expense incurred by a company after it has made a decision to build a plant and has arranged the financing and acquired (or appropriated) the land (if needed) to build it. All of these capital costs are incurred before the plant is taken over by the operating personnel to put it on stream.

Because the costs for some items are largely dependent upon local conditions, they are not included in the cost data reported herein. These include site clearing and preparation, insurance and taxes during construction, and company overhead allocated to the construction project. The estimated costs will vary from system to system. For preliminary estimating purposes the figures shown below represent an average for the recovery alternatives considered in this report.

Land:	Highly site specific. May run approximately \$100,000 per acre and up.
Civil Work and Foundations:	\$45,000, assuming average soil conditions and no major removal of old structures (not included in fluidized bed cost estimate only).
Taxes:	Depends on current legislation. These systems presently fall within a category designated by the Quebec Provincial Government as tax-free since the systems may be considered as environmental improvement.
Company Overhead:	This cost may be included in engineering cost.
Allowance:	A 30 percent allowance is included to cover the following items: - miscellaneous components not identifiable until detailed construction drawings are prepared, - contingencies for price and/or scope changes, strikes, and so on, - engineering and project administration cost.

The estimated accuracy of these preliminary cost estimates should be in the order of plus 20 percent, minus 15 percent accuracy for the items covered in the estimate.

The following information will provide preliminary engineering cost information on some of the liquor recovery systems discussed in earlier sections of this report. It is not intended in the information presented in these hypothetical mill evaluations to directly compare the cost of one process against another. This should be undertaken only

when a site-specific mill application is to be considered. Rather, the information presented herein will focus on costs as they might effect changes in pulping base, pulping yield, and mill size for hypothetical mill conditions representative to the problems of the Canadian sulphite pulping industry now and in the decade ahead.

4.2.1 MEI process liquor recovery cost

The MEI system is of considerable interest in finding a solution to the sulphite liquor problems of Canadian mills. Presently, several Canadian firms and the Pulp and Paper Research Institute of Canada (PPRIC) are evaluating new MEI systems for sodium sulphite, sodium bisulphite, and sodium alkaline sulphite recovery systems. A number of United States installations of the process provide reference information useful in these evaluations. Recent examples include:

- 1) the 70 percent yield sodium bisulphite operation at Georgia Pacific Corporation, Lyons Falls, New York (wherein chemicals are not recycled but sold to kraft mills);
- 2) the ITT Rayonier-Fernandina Beach, Florida installation where hot flue from the ammonia base recovery boiler is used to evaporate hot caustic extraction liquor; and
- 3) a magnesium base recovery system at Nekoosa Papers, Inc., Port Edwards, Wisconsin.

MEI Systems, Inc. believe the magnesium base system generally involves higher capital and operating cost, along with scale and plugging problems in the boilers, evaporators and the absorption system. Thus, MEI does not now offer a system for magnesium base systems and in most of these cases conversion to a sodium base system with full chemical recovery is recommended. This would be the situation in case examples 5 and 6 presented in Table 11.

Table 11 presented the information which was the basis for developing the cost information. Figures 16A and 16B (Section 3) showed the direct three-effect evaporation and burning system which is applied in these evaluations.

The MEI sodium base system, shown in Figure 16B, is applicable to case examples 1, 2, and 3. This does not show the chemical recovery

system. The chemical recovery system does not include the flaker operation. The cost includes stainless steel storage tank requirements including 12 hours of feed liquor and concentrated liquor storage.

The MEI application for ammonia base dissolving pulp is shown in Figure 16A. It is applicable to case examples 4 and 7. For ammonia base NSSC applications a furnace would replace the recovery boiler. These conditions are representative of case examples 4 and 7, respectively, in Table 11. In the case of ammonia base dissolving pulp, the MEI system is compared to a conventional system. Substantial steam is produced in the case of the alternatives for ammonia base dissolving pulp. However, in the conventional ammonia base burning system, considerable steam is consumed (especially for evaporator heating) such that the net steam produced is substantially more for the MEI system. The operating process steam in this case is used for feed water heating, soot blowing, atomizing, liquor heating eductors and, in the case of the conventional system, evaporator heating. Obviously, for all of the case examples, a higher net steam production is possible with higher liquor feed concentration.

Table 12 presents the results of the cost analysis of the case examples for the MEI system. The manpower requirements for all of the recovery systems is two men per shift excepting the case example 7a. Here, conventional recovery requires three men per shift.

It can be noted from Table 12 that the ratio of equipment cost to total installed cost is approximately 0.25 for all of the case examples except 7 and 7a (ammonia base dissolving grade), in which case this ratio increases, reflecting the greater amount of equipment required in the low yield ammonia base operation.

A comparison of installed cost for case examples 1 and 2 shows the effect of economy of scale. The cost of the recovery system per unit of production is less for a large mill than for a small mill. Comparison of case examples 3 and 4 shows that the cost of NSSC and NSSC-NH₃ base is nearly the same for a 250 tpd mill production. Although not clearly shown in these example comparisons, the cost of the recovery system is substantially less for a high yield pulping process than a low yield pulping process.

Separate operating cost data are not reported herein. However, MEI Systems, Inc. claim the return on investment for recovery of chemicals

TABLE 12. SUMMARY OF COST INFORMATION FOR MEI RECOVERY SYSTEMS

Case Example	Alternative	Equipment Cost \$/adt	Installed Cost \$/adt	Power (Kw/adt)	Steam (lb/adt)	Burner	Evaporator	Chemical Recovery
1	NaHSO ₃ (120 t/d)	8,600	34,600	72	670	MEI Smelter	MEI Direct	MEI System
2	NaHSO ₃ (250 t/d)	6,400	25,800	72	670	MEI Smelter	3-Effect MEI Direct	MEI System
3	NSSC (250 t/d)	8,300	33,400	47	435	MEI Smelter	3-Effect MEI Direct	MEI System
4	NSSC-NH ₃ (250 t/d)	8,500	34,000	45	435	Furnace	3-Effect MEI Direct	MEI System
5	MgO (250 t/d)	Recommend Conversion to Sodium Base						
6	MgO (250 t/d)	Recommend Conversion to Sodium Base						
7	NH ₃ (500 t/d) Dissolving	14,100	34,400	86	10,530	Recovery Boiler (1000°F Flue)	MEI Direct 3-Effect	MEI System
8	NH ₃ (500 t/d) Dissolving with Conventional Recovery	16,200	39,500	86	7,200	Recovery Boiler (400°F Flue)	Steam 4-Effect	MEI System

can represent a savings in case example number 2 estimated to run about \$1,300,000 per year.

4.2.2 Fluidized bed liquor recovery cost

The fluidized bed combustion process has been operated since 1973 at the sodium acid sulphite mill of the Ontario Paper Company, Thorold, Ontario. The liquor processed in this case is actually from an alcohol and vanillin plant. Evaporation and recovery of both Na_2SO_4 (64 percent recovery) and Na_2CO_3 (36 percent recovery) is practiced. A fluidized bed recovery system has also been purchased for the new corrugating medium plant of Papier Cascades (Cabana), Inc., Cabana, Quebec. In this case the cooking liquor is prepared by the Owens-Illinois sulphur free process. The Copeland Process is used for both of these fluidized bed recovery processes.

Table 13 presents the technical and cost information for a fluidized bed recovery system considered for the case examples in this report. This cost information is only for the equipment necessary for an operable system and includes fabrication, erection, and construction of the equipment. The cost data do not include civil works, foundations, tanks, and buildings.

The liquor is evaporated by multiple effect evaporators from the dry solids concentration of feed liquor to evaporators indicated in Table 11 up to the necessary dry solids concentration required for self supporting combustion in the fluid bed reactors. The solids concentration is lower in a fluidized bed than in conventional systems and ranges from approximately 30 to 45 percent for the different types of liquors.

For the MgO recovery systems (case examples 5 and 6), a waste heat boiler is included, producing 90,000 and 40,000 lb per hour of steam, respectively. The data shown in the steam column in Table 12 reflects net steam conditions.

Table 14 lists the major equipment items included in the cost analysis.

An analysis of fluidized bed recovery system cost data for case examples 1 and 2 show that the recovery system cost per unit of production is less for a large mill than for a small mill. A similar situation

TABLE 13. SUMMARY OF COST INFORMATION FOR MAJOR EQUIPMENT ITEMS OF FLUIDIZED BED RECOVERY SYSTEM

Case Example	Pulping Alternative	Tons/Day	Yield %	Heat Load	Reactor*	Total	\$ Per Ton Production	Chemical Recovery %	Power KW/ADT	Steam lb/adt	Man-Power	Water gal/adt
1	NaHSO ₃	120	70	26 MM	CRP-10	\$1,400,000	\$11,670	Na - 100 S - 50	75	-7,100	1	16,200
2	NaHSO ₃	250	70	60 MM	CRP-15	2,200,000	8,000	Na - 100 S - 50	40	-4,200	1	9,600
3	NSSC	250	80	39 MM	CRP-12	1,600,000	6,400	Na ₂ SO ₄	40	-4,200	1	9,600
4	NSSC/NH ₃	250	80	50 MM	CRP-12	2,000,000	8,000	S - 95	29	-2,640	1	7,250
5	MgO	250	50	136 MM	CRP-24	4,300,000	17,200	S - 95 MgO - 90	79	+3,500	1	7,250
6	MgO	250	70	62 MM	CRP-19	3,500,000	14,000	S - 95 MgO - 90	46	Even	1	11,600
7	NH ₃ -Dissolving	500	44	420 MM	CRP-38	6,000,000	12,000	S - 92	116	-10,000	1	29,000

* Refers to Copeland Equipment

TABLE 14. LIST OF MAJOR EQUIPMENT ITEMS INCLUDED IN THE FLUIDIZED BED COST ANALYSIS

Fluidized bed reactor	SO ₂ cooling and absorption tower
Fluidizing air blower with drive	Heat exchanger
Air preheater	Pumps and motors
Liquor feed system	Instrumentation including control panel
Refractory	Electricals including motor control center
Bucket elevator	Piping
Product silo	Structural steel
Product metering and cooling screw	Cyclones
Ductwork	Waste heat boiler (MgO systems only)
Venturi evaporator/scrubber	MgO dissolving tank, filter, slaker
Secondary scrubber	Multiple effect evaporator

exists for the ammonia base dissolving pulp alternative. In that case, a 44 percent yield, 100 ton per day mill recovery system would cost approximately \$3,100,000 (\$31,000 per ton), as compared to \$12,000 per ton for the 500 ton per day case example 7 mill.

Cost comparisons for the magnesium base alternatives 5 and 6 reflect the influence of product yield on cost of the recovery system.

The ammonia base NSSC alternative is more expensive per unit of production than the standard NSSC base pulping alternative.

The cost of a fluidized bed recovery system per unit of production is higher for a magnesium base than for a similar sodium bisulphite mill. This reflects the larger reactor required for magnesium base mills.

4.2.3 Cost of liquor recovery with standard set of evaporators and conventional recovery furnace

Liquor recovery cost estimates were also made for a conventional recovery system for case examples 3 through 7. This system consists of a standard set of evaporators and a conventional recovery furnace. A large number of these systems have been installed in North America, beginning in 1948 for the magnesia base alternative [33]. Table 15 presents a summary of the cost information for conventional recovery. The cost of the evaporators required along with this system is itemized separately in Table 16.

These data show that the cost for a conventional recovery system is influenced more by pulping yield than by chemical pulping base. The data also reflect the premium for capital equipment investment for the MgO system per unit of pulp production capacity. This was observed earlier for the fluidized bed recovery system.

4.2.4 Sonoco sulphite recovery process cost

The available cost data for this process are limited mainly to experience based on installation at the Sonoco Products corrugating mill at Hartsville, South Carolina. In that case, a 250 tpd corrugating board mill, operating at about 78 percent liquor recovery, required an estimated total capital cost of \$5,900,000 for evaporation and recovery equipment. This represents \$23,600 per ton of daily production capacity.

TABLE 15. SUMMARY OF COST INFORMATION FOR CONVENTIONAL LIQUID RECOVERY SYSTEM

Case Example	Pulping Alternative	ton/day	Yield %	Equipment Cost, \$			\$ Per Ton Production
				Recovery System	Auxiliary	Total	
3	NSSC	250	80	2,300,000	2,370,000	4,670,000	18,680
4	NSSC-NH ₃	250	80	2,570,000	2,370,000	4,940,000	19,760
5	MgO	250	50	8,500,000	5,000,000	13,500,000	54,000
6	MgO	250	70	6,700,000	4,000,000	10,700,000	42,800
7	NH ₃ -dissolving	500	44	14,900,000	8,950,000	23,850,000	47,700

Equipment Includes:

System

Boiler and Furnace
 Precipitator (98%)
 All Control Systems
 Piping
 Absorption System*
 Acid Equipment*
 Heat Exchanger*
 Mix Tanks**
 Dissolving Tanks**
 Fuel Pumps

Auxiliary

Building
 Service Piping
 Electrical
 Foundations
 Steel
 Stacks

* Not included in Case 3.

** Not included in Cases 4-7.

TABLE 16. SUMMARY OF COST INFORMATION FOR STANDARD SET OF EVAPORATORS

Case Example	Pulping Alternative	ton/day	Yield %	Installed Cost (\$)	\$ Per Ton Production	Description	Required Steam* lb/hr
3	NSSC	250	80	800,000	3,200	Quintuplet	13,400
4	NSSC-NH ₃	250	80	800,000	3,200	Quintuplet	15,450
5	MgO	250	50	3,900,000	15,600	8 Body-5 Effect	39,050
6	MgO	250	70	2,800,000	11,200	8 Body-5 Effect	18,200
7	NH ₃ -dissolving	500	44	6,000,000	12,000	8 Body-5 Effect	112,000

* Assumed steam economy = 5.0

To this must be added approximately 5,000 lb/hr miscellaneous steam for heating the boilers and 18,000 lb/hr for soot blowing. The steam for soot blowing is required only during periods of use.

These production conditions are nearly the same as case example number 3, a NSSC mill, for which the equipment cost was closely comparable to an extrapolation of the cost data presented in Tables 15 and 16 for a standard set of evaporators and a conventional recovery furnace [34].

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APPENDIX

SULPHITE OPERATIONS IN CANADA

APPENDIX

SULPHITE OPERATIONS IN CANADA

A LIST OF NSSC AND SULPHITE MILLS INTEGRATED WITH PAPERMAKING OPERATIONS

<u>MILL NAME AND LOCATION</u>	<u>TYPE (BASE)</u>	<u>YIELD</u>	<u>PRODUCTION (adt/d)</u>
<u>Newfoundland</u>			
Bowater Newfoundland Ltd. Cornerbrooke, Nfld.	HY (Na)	60	330
Price (Nfld.) Pulp and Paper Ltd. Grand Falls, Nfld.	LY (Na)	51	232
<u>New Brunswick</u>			
Fraser Companies Ltd. Edmundston, N.B.	LY (NH ₃) (some market pulp)	46	490
Lake Utopia Paper Ltd. Saint John, N.B.	NSSC (NH ₃)	75-80	200
New Brunswick International Edmundston, N.B.	HY (Na)	63	200
<u>Nova Scotia</u>			
Bowater Mersey Paper Co. Ltd. Liverpool, N.S.	LY (Na)	48	165
Nova Scotia Forest Industries Ltd. Port Hawkesbury, N.S.	HY (Na)	67	110
<u>Quebec</u>			
QNS Paper Company Ltd. Baie Comeau, P.Q.	HY (Na)	64	340
Consolidated Bathurst Ltd. Shawinigan, P.Q.	HY (Na)	72	265
Consolidated Bathurst Ltd. Port Alfred, P.Q.	HY (Na)	73	215
The Price Company Ltd. Alma, P.Q.	HY (Na)	70	200

The Price Company Ltd. Kenogami, P.Q.	HY (Na)	68	200
Abitibi Paper Co. Ltd. Beaupre, P.Q.	HY (Na)	70	105
Canadian International Paper Co. Three Rivers, P.Q.	LY (Na)	47	260
The Price Company Ltd. Chandler, P.Q.	LY (Na)	52	175
The Donohue Company Ltd. Cleremont, P.Q.	LY (Na)	52	160
Consolidated Bathurst Ltd. Grand'Mere, P.Q.	LY (Na)	50	130
Three Rivers Pulp & Paper Co. Ltd. Three Rivers, P.Q.	LY (Na)	50-54	102
Papeterie Reed Ltée Quebec Mill Division Quebec, P.Q.	LY (Ca)	47	280
Domtar Newsprint Ltd. Dolbeau, P.Q.	LY (Ca)	48	175
Domtar Newsprint Ltd. Donnacona, P.Q.	LY (Ca)	49	127
The James MacLaren Co. Ltd. Masson, P.Q.	LY (Mg)	51	100
Canadian International Paper Co. Matane, P.Q.	NSSC (Na)	72	162
Domtar Packaging Ltd. East Angus, P.Q.	NSSC (Na)	75	108

Ontario

Abitibi Paper Co. Ltd. Iroquois Falls, Ont.	HY (Na)	65	264
Abitibi Paper Co. Ltd. Thunder Bay, Ont.	HY (Mg)	64-66	110
Abitibi Paper Co. Ltd. Fort William, Ont.	HY (Mg)	64	82
The Ontario Paper Co. Ltd. Thorold, Ont.	LY (Na)	48	240

Ontario-Minnesota Pulp & Paper Co. Ltd. Kenora, Ont.	HY (Mg)	63	210
Abitibi Provincial Paper Co. Ltd. Thunder Bay, Ont.	LY (Ca)	50	95
The Great Lakes Paper Co. Ltd. Thunder Bay, Ont.	LY (Mg)	50	350
Spruce Falls Power & Paper Co. Ltd. Kapuskasing, Ont.	LY (Ca) LY (Mg)	44 50	350 120
Abitibi Forest Products Ltd. Sturgeon Falls, Ont.	NSSC (Na)	80	242

Manitoba

Abitibi Paper Co. Ltd. Pine Falls, Man.	HY (Na)	65	105
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B LIST OF MARKET SULPHITE PULP MILLS

Fraser Companies, Ltd. Atholville, N.B.	LY (NH ₃)	46	284
Irving Pulp and Paper Ltd. St. John, N.B.	LY (NH ₃)	46	200
Nova Scotia Forest Industries Ltd. Port Hawkesbury, N.S.	LY (Na)	46	450
Rayonier Quebec Inc. Port Cartier, P.Q.	LY (NH ₃)	44	750
Tembec Forest Products Inc. Temiskaming, P.Q.	LY (NH ₃)	46	450
St. Raymond Paper Ltd. Desbiens, P.Q.	LY (Mg)	49	130
Canadian Intern. Paper Co. Ltd. Hawkesbury, Ont.	LY (NH ₃) (Dissolving Pulp)	43	287
Rayonier Canada (B.C.) Ltd. Port Alice, B.C.	LY (NH ₃) (Dissolving Pulp)	44	450

C LIST OF SULPHITE PLANTS THAT HAVE DISCONTINUED OPERATIONS IN THE LAST 10 YEARS

1. Abitibi Paper Company Ltd., Sault St. Marie, Ont.
2. Canadian Cellulose Co. Ltd., Prince Rupert, B.C.

3. Canadian International Paper Co., Gatineau, P.Q.
4. Domtar Fine Papers Ltd., Cornwall, Ont.
5. The E.B. Eddy Company, Hull, Quebec.
6. Kruger Pulp and Paper Ltd., Bromptonville, P.Q.

D LIST OF SULPHITE MILLS THAT EVAPORATE, INCINERATE, AND RECOVER
SOME COOKING CHEMICALS

1. Nova Scotia Forest Industries Ltd., Port Hawkesbury, N.S.
2. Rayonier Quebec Inc., Port Cartier, P.Q.
3. The Ontario Paper Co. Ltd., Thorold, Ont.
4. Spruce Falls Power and Paper Co. Ltd., Kapuskasing, Ont.
5. Papeterie Reed Ltée, Quebec, Quebec

Nova Scotia Forest Products Ltd. - Port Hawkesbury, Ontario

This company used the basic Stora liquor recovery process whereby elemental sulphur and a sulphite/bisulphite mixture is recovered for use in the pulping operations. A part of the above mixture is fortified with additional sulphur dioxide to produce the bisulphite cooking liquor for the high yield sulphite plant.

Rayonier Quebec Inc. - Port Cartier, P.Q.

This mill recovers sulphur dioxide only by stepwise concentration by evaporation and burning in a recovery boiler. The flue gas is passed through a flue gas scrubber where anhydrous ammonia and water are used to produce the ammonia base bisulphite liquor.

The Ontario Paper Company Ltd. - Thorold, Ontario

This mill has installed the Copeland Fluidized Bed Recover System wherein some of the spent sulphite liquor and the liquid wastes from the alcohol and vanillin by-products are treated. Recovered products are sodium sulphate and carbonate. The salt cake is sold on the market. Water vapor, nitrogen and carbon dioxide are discharged to the atmosphere.

Spruce Falls Power and Paper Company Ltd. - Kapuskasing, Ontario

This mill evaporates, incinerates, and recovers magnesium and sulphur dioxide cooking chemicals from the higher yield pulping operations only (Magnefite Process).

Papeterie Reed Ltée - Quebec, Quebec

This mill collects and evaporates approximately one-third of their spent liquor to produce byproducts.

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